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Impact of Ammonia Utilization by NOx Flue Gas Treatment Processes

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16. ABSTRACT
The report gives results of a study of the impact of ammonia (NH3) utilization by NOx flue gas treatment (FGT) processes. The most technologically advanced FGT system for the highly efficient (about 90%) removal of NOx from power plant stack gas is selective catalytic reduction (SCR) using NH3. A major economic consideration in the widespread application of SCR is the impact of this demand on the domestic NH3 market. Annual NH3 requirements for NOx FGT control of new, fossil-fuel-fired boilers were calculated and predicted for the period 1978-2000. The total NH3 supply was also projected for the same period and compared with the projected annual demand for NOx FGT. The study concludes that NH3 demand for NOx FGT systems will begin gradually and rise at relatively slow rates, giving the domestic NH3 market adequate time to adjust to the increased demand. Under other than study assumptions (e.g., requiring retrofit of NOx FGT systems on all large boilers), significant impacts and market disruptions are foreseen. The current and projected NH3 production techniques were outlined and the 1978 unit production cost for NH3 was estimated for four potential feedstocks.

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Flue Gases	Combustion	Stationary Sources	21B	
Gas Purification		Flue Gas Treatment	13H,07A	
Ammonia		Selective Catalytic Reduction	07B	
Nitrogen Oxides				
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ABSTRACT

At the present time, the most technologically advanced type of flue gas treatment (FGT) system for the removal of nitrogen oxides (NO_x) from power plant stack gas is selective catalytic reduction (SCR) using ammonia (NH_3). One of the major economic considerations in the widespread application of these FGT systems is the impact of this demand on the domestic NH_3 market. Annual NH_3 FGT requirements for typical fossil fuel-fired boilers were calculated and the annual NH_3 requirement for NO_x FGT was predicted for the period 1978-2000. The total NH_3 supply and demand were also projected for the period 1978-2000 and compared with the projected annual NH_3 demand for NO_x FGT. Based on the premises of this study this NH_3 demand for NO_x FGT systems would begin gradually and rise at relatively slow rates, thus giving the domestic NH_3 market adequate time to adjust to this increased demand. Under other assumptions, such as requiring NO_x FGT systems on all large boilers, significant impacts and market disruptions can be foreseen. The current and projected NH_3 production techniques were outlined and the 1978 unit production cost for NH_3 was estimated for four potential feedstocks. Since the hydrogen (H_2) feedstock cost represents a significant portion of the cost of generating NH_3 and the cost of these feedstocks during the period 1978-2000 is uncertain, the unit production costs for NH_3 during the period 1980-2000 were not estimated.

CONTENTS

Abstract ii
Figures v
Tables vii
Abbreviations and Conversion Factors x

Executive Summary xi

Introduction 1
  Formation of NO_x in Large Boilers 3
  Potential NO_x Control Methods for Large Boilers 3

NH_3 Requirements for NO_x Control 7
  NH_3 Requirements for Typical Power Plant Applications 7
  Projection of Coal-Fired Generating Units 12
    NH_3 Requirements for Both Utility and Industrial Boilers 19
  Current NH_3 Supply and Demand 22
  Comparison of Projected NH_3 Demand for NO_x FGT with Potential NH_3 Supply 23

Current NH_3 Generating Techniques 33
  Steam Reforming of Natural Gas 33
    Processing Scheme 34
    CO Conversion 34
    Final Purification 38
  NH_3 Synthesis 39
    Raw Material, Utility, Labor, and Energy Requirements 41
    Process Economics 41
  Other Current NH_3 Generating Techniques 43
    Refinery Gas 47
    Electrolytic Cells 48
    Coke-Oven Gas 48
    Raw Material, Utility, Labor, and Energy Requirements 51

Potential Future NH_3 Generating Techniques 52
  Steam Reforming of Naphtha 52
    Process Description 52
    Raw Material, Utility, Labor, and Energy Requirements 54
    Process Economics 54
  Partial Oxidation of Hydrocarbons 55
    Process Description 60
    Raw Material, Utility, Labor, and Energy Requirements 64
    Process Economics 64

NH_3 from Coal 65
  Process Description 65
  Raw Material, Utility, Labor, and Energy Requirements 74
  Process Economics 74
NH_3 Imports 76
  Projected Mexican NH_3 Capacity and Production 76

Economic Comparison of NH_3 Production Cost for Various H_2 Feedstocks. 81

Current and Projected Natural Gas Supply and Demand 83
  Conventional Natural Gas Supplies 86
  Supplemental Natural Gas Supplies 87
    SNG from Oil and Coal 87
    Importation of Natural Gas from Canada 89
    Importation of Natural Gas from Mexico 91
    Alaskan Natural Gas 93
    Importation of Liquefied Natural Gas 93
    New Technologies 98
    Other Sources of Natural Gas 99

Conclusions 101
  Onsite Generation Versus Onsite Storage of NH_3 for NO_x FGT Systems. 101
  Impact of NH_3 Demand for NO_x FGT on the Domestic NH_3 Market 101
  Current NH_3 Production Economics 103

References 104

FIGURES

Number **Page**

S-1 Projected change in emissions for the five major air pollutants during the period 1975-1990 xvi
S-2 Comparison of the projected NH_3 demand for NO_x FGT with conventional NH_3 demand for selected years during the period 1980-2000 (case 1 only) xxii
1 Breakdown of the total U.S. NO_x emissions by original source 2
2 NO_x removal efficiency as a function of NH_3:NO_x mol ratio. . 9
3 Annual NH_3 consumption for selected 500-MW boilers for NO_x FGT as a function of NH_3:NO_x mol ratio 10
4 U.S. electrical generating capacity by fuel type for the years 1975-1985 14
5 Projected U.S. electrical generating capacity for the period 1985-2000 at three different annual growth rates . . 15
6 Projected U.S. NH_3 demand during the period 1975-2000 at three annual growth rates 24
7 Annual operating rate of U.S. NH_3 plants from 1965-1981 . . 26
8 Projected U.S. NH_3 capacity during the period 1975-2000 at three annual growth rates 27
9 Comparison of the projected NH_3 demand for NO_x FGT with conventional NH_3 demand for selected years during the period 1980-2000 (case 1 only) 29
10 Block flow diagram of a typical NH_3 plant based on natural gas as the H_2 feedstock 35
11 Diagram of typical primary reformer 36
12 Diagram of quench-type converter 40
13 Block flow diagram of a typical NH_3 plant based on coke-oven gas as the H_2 feedstock 50
14 Block flow diagram of a typical NH_3 plant based on naphtha as the H_2 feedstock 53
15 Block flow diagram of a typical NH_3 plant based on heavier hydrocarbons as the H_2 feedstock 59
16 Block flow diagram of a typical NH_3 plant based on coal as the H_2 feedstock 69
17 Net U.S. imports and exports of NH_3 during the period 1965-1981 79
18 Geographic location of U.S. NH_3 plants operating before 1975 84
19 Geographic location of new U.S. NH_3 plants operating after 1975 85

FIGURES (continued)

Number Page

20 Proven U.S. natural gas reserves during the period 1968-1980 88

21 Natural gas imports from Canada during the period 1955-1976 90

22 Natural gas imports from Mexico during the period 1955-1976 92

TABLES

Number Page

S-1 NH_3 Consumption for Typical 500-MW Coal- and Oil-Fired Boilers xvii
S-2 Projected U.S. Electrical Generation Capacity by Fuel for Selected Years During the Period 1975-2000 xviii
S-3 Projected Coal-Fired Power Plant Additions and the Resulting NH_3 Demand for NO_x FGT During the Period 1984-2000 xix
S-4 Comparison of Annual NH_3 Demand for NO_x FGT for Three Options During the Period 1985-2000 xxi
S-5 Comparison of Unit Labor and Energy Requirements per Ton of NH_3 Produced from Various Potential H_2 Feedstocks in 1978 xxiii
S-6 Alternative Feedstock Costs and the Resulting Unit Production Costs of NH_3 in 1978 xxiv
1 NO_x Emissions by Source in 1975 3
2 Effects of Boiler Modifications to Reduce NO_x Emissions by Fossil Fuel Type 5
3 Flue Gas Compositions from 500-MW Coal- and Oil-Fired Boilers 8
4 Premises for the Calculation of Annual NH_3 Consumption for 500-MW Fossil-Fired Boilers 11
5 NH_3 Consumption for Typical 500-MW Coal- and Oil-Fired Boilers 11
6 Projected U.S. Electrical Generating Capacity by Fuel Type for the Years 1975-1985 12
7 Estimated U.S. Generating Capacity Additions by Fuel Type for the Period 1976-1985 13
8 Projected Electrical Generating Capacity by Fuel Type in the Year 2000 16
9 Projected Coal-Fired Power Plant Additions and the Total Coal-Fired Generating Capacity by Year for the Period 1984-2000 17
10 Comparison of Projected Annual U.S. NH_3 Consumption by Alternative NO_x FGT Systems at all New Coal-Fired Power Plants 18
11 Estimated NO_x Emissions from Various Sized Boilers in 1975 and 1985 20
12 Comparison of the Projected Annual NH_3 Demand for NO_x FGT During the Period 1985-2000 for Two Cases: Required on Only New Utility Boilers or for All New Large Boilers 21

TABLES (continued)

Number		Page
13	U.S. NH_3 Demand 1965-76	22
14	U.S. NH_3 Plant Capacity Additions and Total U.S. NH_3 Plant Capacity During the Period 1967-1980	25
15	Comparison of the Estimated Annual NH_3 Demand for NO_x FGT with the Projected Conventional U.S. NH_3 Demand During the Period 1985-2000	28
16	Projected U.S. NH_3 Demand for NO_x FGT for the Period 1985-2000 if all Coal-Fired Power Plants Required Treatment	31
17	Typical Natural Gas Composition	33
18	Typical Gas Composition Entering and Leaving Secondary Reformer	37
19	Typical Synthesis Gas Composition Before and After Methanation	39
20	Unit Raw Material, Utility, Labor, and Energy Requirements for the Production of NH_3 from Natural Gas	41
21	Annual Capital Charges for Chemical Industry Financing	42
22	Assumptions Used to calculate the NH_3 Production Costs	43
23	Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Natural Gas as the H_2 Feedstock (Low Feedstock Cost Case)	44
24	Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Natural Gas as the H_2 Feedstock (Medium Feedstock Cost Case)	45
25	Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Natural Gas as the H_2 Feedstock (High Feedstock Cost Case)	46
26	Typical Compositions of Refinery Off-Gases	47
27	Typical Composition of Coke-Oven Gas	49
28	Unit Raw Material, Utility, Labor and Energy Requirements for the Production of NH_3 from Coke-Oven Gas	51
29	Unit Raw Material, Utility, Labor, and Energy Requirements for the Production of NH_3 from Naphtha by Steam Reforming	54
30	Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Naphtha (Steam Reforming) as the H_2 Feedstock (Low Feedstock Cost Case)	56
31	Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Naphtha (Steam Reforming) as the H_2 Feedstock (Medium Feedstock Cost Case)	57
32	Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Naphtha (Steam Reforming) as the H_2 Feedstock (High Feedstock Cost Case)	58
33	Typical Operating Parameters for Production of NH_3 Synthesis Gas from Various Hydrogen Feedstocks Using the Shell Gasification (Partial Oxidation) Process	61

TABLES (continued)

Number Page

34 Typical Feedstocks Used in the Shell Gasification Process (Partial Oxidation) . . . 62
35 Typical Compositions of Gases to and from a Cryogenic Purification Unit . . . 63
36 Unit Raw Material, Utility, Labor, and Energy Requirements for the Production of NH_3 from Heavy Fuel Oil . . . 64
37 Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Heavy Fuel Oil as the H_2 Feedstock (Low Feedstock Cost Case) . . . 66
38 Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Heavy Fuel Oil as the H_2 Feedstock (Medium Feedstock Cost Case) . . . 67
39 Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Heavy Fuel Oil as the H_2 Feedstock (High Feedstock Cost Case) . . . 68
40 Typical Exit Gas Compositions for Low and High Temperature Gasifiers . . . 71
41 Typical Synthesis Gas Composition Before and After Co Shift . . . 72
42 Typical Synthesis Gas Compositions After CO_2 Wash and After Liquid N_2 Wash . . . 73
43 Comparison of Typical Raw Gas Composition from the Texaco With That for a High Temperature Gasifier . . . 73
44 Unit Raw Material, Utility, Labor, and Energy Requirements for the Production of NH_3 from Coal . . . 74
45 Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Coal as the H_2 Feedstock (Low Feedstock Cost Case) . . . 75
46 Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Coal as the H_2 Feedstock (Medium Feedstock Cost Case) . . . 77
47 Total Annual Operating Costs and Unit Operating Costs for a 1000 Ton/Day NH_3 Plant Using Coal as the H_2 Feedstock (High Feedstock Cost Case) . . . 78
48 Projected Mexican NH_3 Plants . . . 80
49 Alternative Feedstock Costs and the Resulting Unit Production Cost of NH_3 in 1978 . . . 82
50 Estimates of U.S. Natural Gas Reserves . . . 87
51 Natural Gas Reserves of Various Countries . . . 94
52 LNG Exporting Countries in 1985 . . . 95
53 Current and Proposed LNG Base-Load Plants . . . 96
54 Current and Projected LNG Imports and LNG Exports by Country . . . 97

ABBREVIATIONS AND CONVERSION FACTORS

ABBREVIATIONS

aft^3	actual cubic feet	lb	pound
Btu	British thermal unit	LNG	liquefied natural gas
^{o}C	degrees Centigrade	M	million (mega = 10^6)
ESP	electrostatic precipitator	Mft^3/day	million cubic feet per day
^{o}F	degrees Fahrenheit	mol	mole
FGD	flue gas desulfurization	MW	megawatt
FGT	flue gas treatment (NO$_x$)	MWe	megawatt electric
ft	feet	NGL	natural gas liquids
ft^3	cubic feet	Nm^3	normal cubic meter
gal	gallon	NO$_x$	nitrogen oxides
G	billion (giga = 10^9)	ppm	parts per million
Gft^3	billion cubic feet	psig	pounds per square inch gauge
GW	gigawatt	sft^3	standard cubic feet
hr	hour	SCR	selective catalytic reduction
IC	internal combustion	sec	second
k	thousand	SNG	synthetic natural gas
kW	kilowatt	T	trillion (tera = 10^{12})
kWh	kilowatthour	Tft^3	trillion cubic feet
		yr	year

CONVERSION FACTORS

To convert from	To	Multiply by
British thermal unit	gram-calories	252
degrees Fahrenheit-32	degrees Centigrade	0.5555
feet	centimeters	30.48
cubic feet	cubic meters	0.02832
feet per minute	centimeters per second	0.508
cubic feet per minute	cubic meters per second	0.000472
gallons	liters	3.785
gallons per minute	liters per second	0.06308
grains per cubic foot	grams per cubic meters	2.228
megawatt (electric)	standard cubic feet per minute	0.000535
pounds	kilograms	0.4536
tons (short)	metric tons	0.90718
tons per hour	kilograms per second	0.252

THE IMPACT OF AMMONIA UTILIZATION BY NO_x FLUE GAS TREATMENT PROCESSES ON THE DOMESTIC AMMONIA MARKET

EXECUTIVE SUMMARY

Although nitrogen oxide (NO_x) concentration in each individual emission source is relatively low, the number of these potential sources is increasing and thus the total amount of NO_x emitted in the United States is projected to increase more than 50% during the period 1975-1990 under the current NO_x control regulations. Even with the widespread application of the best control technology on all new sources, the total U.S. NO_x emissions are expected to increase by about 25% during this same time period. Since NO_x is the only one of the five major air pollutants to show this dramatic increase (see Figure S-1), the NO_x emission regulations for stationary sources may become more stringent in the future. The potential for additional regulatory control of NO_x emissions from stationary sources is further increased by the recent postponement of NO_x emission regulations on mobile sources.

In terms of total NO_x emissions, large fossil-fueled boilers represent the largest class of stationary sources, releasing over 4.7 Mtons of NO_x in 1975. At the present time, two differing types of NO_x control methods for these large boilers are undergoing development in the United States and Japan. Combustion modifications, which seek to reduce NO_x emissions by controlling boiler operating conditions, and hence preventing the formation of NO_x, are receiving considerable attention in the United States. Although these combustion modifications are relatively inexpensive when compared with other alternatives, they are apparently limited to reducing NO_x emissions by about 50%. Flue gas treatment (FGT) systems which are capable of removing up to 90-95% of the NO_x emitted are receiving the major emphasis in Japan, primarily because the current NO_x emission regulations are much more stringent in Japan. As their name implies these systems involve treating and removing the NO_x from the flue gas after it has been formed. Thus if 90% NO_x control from all large boilers is required in the future, two alternative cases are possible: (1) 90% NO_x removal by FGT or (2) combination of 50% NO_x reduction by combustion modification and 80% removal of the remaining NO_x by FGT. Although case 2 (50% reduction by combustion modification and 80% removal by FGT) would appear to be more attractive if 90% NO_x control is required, no comparative economics have yet been completed and thus both cases will be considered.

From the standpoint of development status, the most advanced FGT systems now are the selective catalytic reduction (SCR) processes. These processes are based on ammonia (NH_3) selectively reducing NO_x in the presence of a base-metal catalyst. A major concern associated with these processes, if widely applied in the United States, is the direct impact of this additional NH_3 demand on the NH_3 market and the subsequent indirect impact on natural gas availability for producing the NH_3.

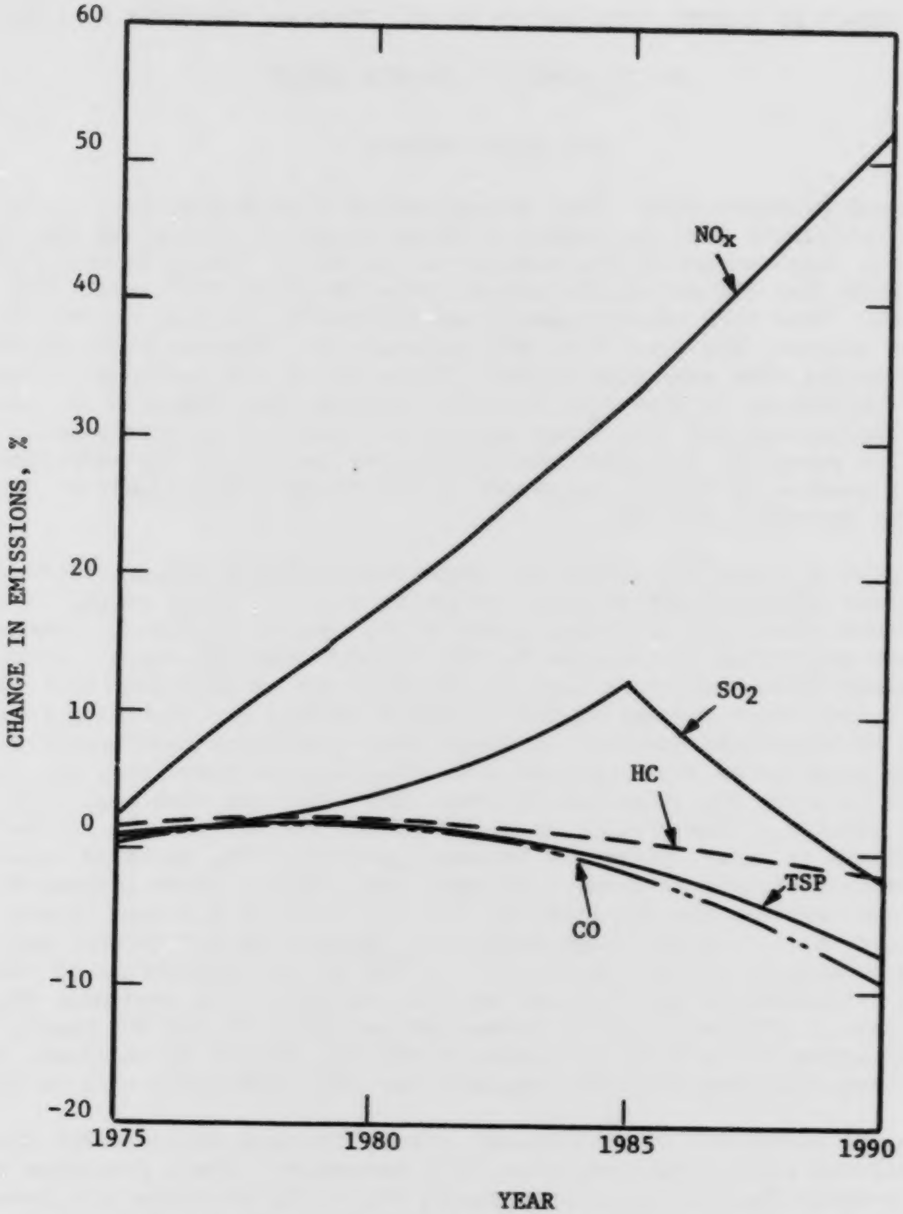

Figure S-1. Projected change in emissions for the five major air pollutants during the period 1975-1990.

NH_3 DEMAND FOR NO_x FGT

The annual NH_3 requirements for 500-MW coal- and oil-fired boilers were calculated for both cases, i.e., SCR alone and combustion modification in combination with SCR. Since the NO_x removal efficiency for an SCR system can best be altered by changing the NH_3:NO_x mol ratio, a generic curve was drawn showing the relationship between the NH_3:NO_x mol ratio and the NO_x removal efficiency. For 90% NO_x removal and 80% NO_x removal (case 1 and case 2), the NH_3:NO_x mol ratios were 1.05:1 and 0.91:1 respectively. As shown in Table S-1, the resulting NH_3 consumption rates for a 500-MW coal-fired power plant range from nearly 6300 tons/yr for the 90% efficient SCR system to about 2700 tons/yr for the combined combustion modification-SCR treatment scheme. For a 500-MW oil-fired power plant the annual NH_3 requirements would range from nearly 1800 tons for the 90% efficient SCR system to about 800 tons for the combined treatment scheme. Thus even a 2,500-MW coal-fired power plant requiring about 31,000 tons/yr could not justify a captive NH_3 plant since the minimum plant capacity for economic production of NH_3 is 1,000 tons/day (330,000 tons/yr). The NH_3 for NO_x control system would therefore probably be purchased from an existing NH_3 plant and shipped to storage tanks at the power plant.

TABLE S-1. NH_3 CONSUMPTION FOR TYPICAL

500-MW COAL- AND OIL-FIRED BOILERS

Fuel	NO_x treatment scheme	NO_x concentration in flue gas, ppm	Mols NH_3 per mol NO_x	NH_3 consumption, tons/yr
Coal	SCR[a]	600	1.05	6,266
Coal	CM & SCR[b]	300[c]	0.91	2,715
Oil	SCR	200	1.05	1,781
Oil	CM & SCR	100[c]	0.91	772

[a] Selective catalytic reduction.

[b] Combustion modification followed by selective catalytic reduction.

[c] Combustion modification was assumed to be 50% efficient in controlling NO_x emissions.

The projected U.S. electrical capacity and the fuel mix in these power plants for the period 1975-2000 was estimated as shown in Table S-2. (Although the specific generating capacity of both oil- and gas-fired boilers in the year 2000 has not been given, no new gas-fired boilers are expected to be built after 1980 and a phaseout of oil-fired boilers is also expected during the period 1985-2000.) By assuming a retirement factor of 0.5% of the total

U.S. installed generating capacity each year, the additional new coal-fired capacity needed each year was calculated. This post-1985 coal-fired capacity then is the electrical capacity potentially required to achieve a 90% NO_x removal. Since the annual NH_3 requirement for a typical 500-MW coal-fired boiler had been previously calculated, the estimated additional NH_3 demand for NO_x FGT could be projected for both cases, i.e., 90% removal by SCR and 80% removal by SCR, as shown in Table S-3. Although the widespread application of NO_x FGT would require substantial amounts of NH_3, the initial increases in demand would be small enough to minimize any adverse impacts on the NH_3 industry, i.e., requiring only one new NH_3 plant to supply the needs of all the NO_x FGT systems that might come online in 1985 and 1986.

TABLE S-2. PROJECTED U.S. ELECTRICAL

GENERATION CAPACITY BY FUEL FOR SELECTED YEARS

DURING THE PERICD 1975-2000

	Electrical generating capacity, GW			
Fuel	1975	1980	1985	2000
Coal	183	247	321	829
Oil	127	155	163	-[a]
Gas	63	59	49	-[a]
Nuclear	33	78	174	536
Hydro[b]	69	80	92	-[a]
Total	474	619	798	1,658

a. Total of 256 GW but not broken down into oil, gas, and hydro although most are expected to be hydro or new technology such as geothermal, solar, etc.

b. Includes hydro, pump storage, and other miscellaneous.

An alternative case would be the setting of standards which require 90% NO_x removal from all new large boilers (i.e., greater than 250 MBtu/hr) rather than only new utility boilers. Detailed data on the projected capacity of new large industrial boilers were not available. Recent historical data, indicating large industrial boilers emit about 25% of all the NO_x emitted from all large boilers, were used to project the additional NH_3 demand for the widespread application of NH_3-based NO_x FGT system for all new large boilers. The NH_3 demand for NO_x FGT under this alternative case then was estimated as

TABLE S-3. PROJECTED COAL-FIRED POWER PLANT ADDITIONS AND THE RESULTING NH_3 DEMAND FOR NO_x FGT DURING THE PERIOD 1984-2000

Year	New coal-fired capacity, GW	Annual NH_3 consumption,[a] ktons, Case 1[b]	Case 2[c]
1984	19	0	0
1985	15	94	41
1986	23	332	144
1987	24	626	271
1988	26	939	407
1989	27	1,271	551
1990	29	1,622	703
1991	31	1,998	866
1992	33	2,399	1,039
1993	35	2,825	1,224
1994	37	3,276	1,419
1995	40	3,759	1,629
1996	42	4,273	1,851
1997	45	4,818	2,088
1998	48	5,401	2,340
1999	51	6,022	2,609
2000	55	6,687	2,897

a. Basis
NO_x FGT required only for new, coal-fired power units coming online after January 1, 1985.

b. Ninety percent removal by selective catalytic reduction.

c. Ninety percent removal by combined combustion modification and selective catalytic reduction.

25% higher than that for new utility boilers only. Although the actual numbers change, the result remains the same, i.e., the initial increases in NH_3 demand are sufficiently small so that no abrupt adverse impacts on the NH_3 industry are anticipated.

The available NH_3 supply to meet both the conventional NH_3 demands and annual growth of this additional demand for NO_x FGT was projected using an assumed annual growth rate of 3.0%. Since most of the current NH_3 demand is for agricultural uses and no substantial increase in the application rate is anticipated, the expected growth rate of the domestic NH_3 demand was projected at slightly below the recent historical rate. Under this scenario present NH_3 production of 16.4 Mtons (1976) would increase to about 33.1 Mtons in the year 2000. A comparison of this projected NH_3 supply with the increased potential demand for NH_3 for NO_x FGT is shown in Table S-4 and Figure S-2 for three alternative scenarios: (1) new large utility boilers only, (2) all new large boilers, and (3) all large boilers, new and old. For each of these scenarios, the NH_3 consumption rate was calculated assuming that a 90% efficient SCR system was used for NO_x FGT. When using combustion modifications in combination with 80% efficient SCR system, the annual NH_3 consumption would be about 57% less. For the first two scenarios, only new large utility boilers and all new large boilers, the additional NH_3 demand for NO_x FGT as a percent of the total NH_3 supply ranges from less than 0.6% in 1985 to about 20-25% in the year 2000. Thus, although requiring substantial amounts of NH_3 during the later years, the initial need for NH_3 for NO_x FGT would not have an immediate substantial impact on the NH_3 market.

The third scenario was included to demonstrate the importance of the premises selected for this study. In an extreme condition such that all large boilers, both new and old, will require 90% NO_x control, the impact of this additional demand in 1985 would create distortions in the NH_3 market with the resulting adverse impact on the fertilizer industry.

CURRENT AND POTENTIAL NH_3 GENERATING TECHNIQUES

Most of the NH_3 produced at the present time is based on steam reforming of natural gas. Although any hydrocarbon could be used, natural gas has the advantages of a high hydrogen to carbon ratio and a high purity as received at the plant. The NH_3 which is produced from H_2 generated by other methods, such as electrolytic chlorine (Cl_2) cells, refinery off-gases, and coke-oven gas, represents <5% of the total U.S. NH_3 supply. Since this H_2 is typically a byproduct stream, it is a relatively cheap source of feedstock but its potential as a future alternative source of NH_3 is severely limited by the demand for the primary products, i.e., Cl_2 and coke.

Other potential feedstocks include petroleum fractions, such as naphtha and fuel oil, as well as coal. The two major problems with these alternate feedstocks are a lower hydrogen to carbon ratio and also higher levels of various impurities. These disadvantages increase the amount of feedstock (on a Btu basis) needed to generate the same amount of H_2 and also increase the capital investment by requiring both pretreatment of the feedstock and the

TABLE S-4. COMPARISON OF ANNUAL NH_3 DEMAND FOR NO_x FGT FOR THREE OPTIONS DURING THE PERIOD 1985-2000

Year	Projected annual NH_3 supply, Mtons	Annual NO_x FGT NH_3 demand,[a] Mtons					
		New large utility boilers only	% of total supply	All new large boilers	% of total supply	All large boilers	% of total supply
1985	21.4	0.09	0.4	0.12	0.6	5.03	23.5
1986	22.1	0.33	1.5	0.42	1.9	5.36	24.3
1987	22.7	0.63	2.8	0.78	3.4	5.70	25.1
1988	23.4	0.94	4.1	1.17	5.2	6.08	26.8
1989	24.1	1.27	5.3	1.59	6.6	6.48	26.9
1990	24.8	1.62	6.5	2.03	8.2	6.89	27.8
1991	25.6	2.00	7.8	2.50	9.8	7.35	28.7
1992	26.4	2.40	9.1	3.00	11.4	7.84	29.7
1993	26.9	2.83	10.5	3.53	13.1	8.34	31.0
1994	27.7	3.28	11.8	4.10	14.8	8.89	32.1
1995	28.5	3.76	13.2	4.70	16.5	9.46	33.2
1996	29.4	4.27	14.5	5.34	18.2	10.09	34.3
1997	30.3	4.82	15.9	6.02	19.9	10.75	35.5
1998	31.2	5.40	17.3	6.75	21.6	11.44	36.7
1999	32.1	6.02	18.8	7.53	23.5	12.19	38.0
2000	33.1	6.69	20.2	8.36	25.3	12.99	39.2

a. Basis

FGT system, 90% NO_x removal by SCR.

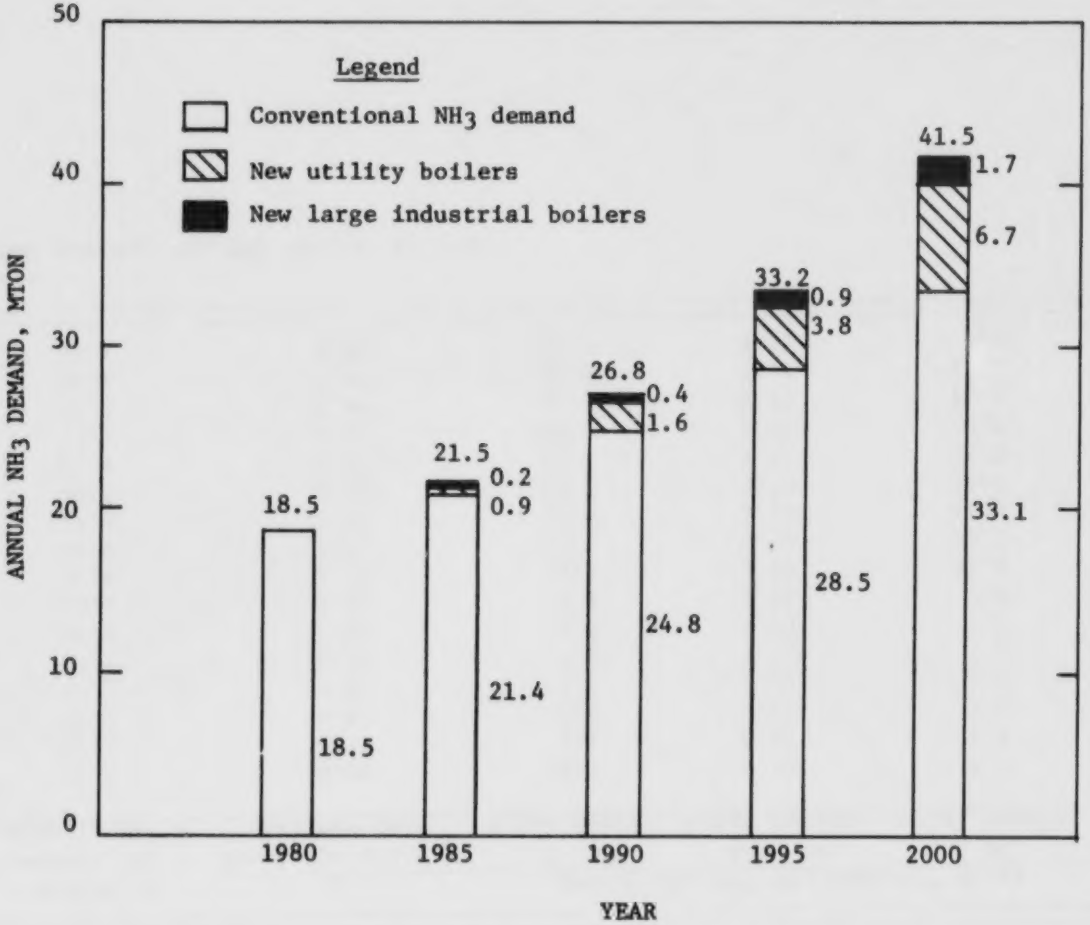

Figure S-2. Comparison of the projected NH_3 demand for NO_x FGT with conventional NH_3 demand for selected years during the period 1980-2000 (case 1 only).

removal of the additional amounts of impurities. Table S-5 provides a comparison of various important operating parameters for generating 1 ton of NH_3 from each of the potential H_2 feedstocks. These unit quantities are based on a 1000 ton/day NH_3 plant and include: feedstock, fuel, other energy, total energy, and labor requirements. As would be expected the labor and the total energy required per ton of NH_3 increases as the feedstock changes from lower molecular weight, less complex hydrocarbon (natural gas) to the higher molecular weight, higher impurity feedstock (coal). In fact the labor requirements to produce 1 ton of NH_3 increases by a factor of 2.5 in going from natural gas to coal as the feedstock while the total energy per ton increases from 31.7 MBtu for natural gas to 41.6 MBtu for coal.

TABLE S-5. COMPARISON OF UNIT LABOR AND ENERGY REQUIREMENTS PER TON OF NH_3 PRODUCED FROM VARIOUS POTENTIAL H_2 FEEDSTOCKS IN 1978[a]

	Natural gas	Naphtha	Fuel oil	Coal
Labor, man-hr	0.18	0.20	0.23	0.45
Energy, MBtu				
Feedstock	19.0	19.4	20.3	27.4
Fuel	12.6	14.1	14.6	13.7
Other	0.1	0.2	0.5	0.5
Total energy	31.7	33.7	35.4	41.6

a. Basis
Plant capacity, 1000 tons/day.
On-stream time, 330 days/yr.

Current NH_3 Production Costs for Various H_2 Feedstocks

The feedstock which will be used for NH_3 production in the future will be determined by process economics and the future prices of these various H_2 feedstocks are uncertain at the present time. Current unit production costs of NH_3 for each of the four potential H_2 feedstocks were estimated using three alternative prices as shown in Table S-6. With the current purchase price of natural gas varying sharply and dependent on numerous factors, an average price for natural gas would be difficult to identify. Therefore the range of prices for natural gas shown in Table S-6 was assumed. Of the three prices assumed for the other three alternate feedstocks (naphtha, fuel oil, and coal), the lower price was typically the current feedstock cost, the medium price was included to reasonably cover the potential range of the current feedstock cost, while the higher cost was included to identify an extreme condition.

TABLE S-6. ALTERNATIVE FEEDSTOCK COSTS AND THE RESULTING

UNIT PRODUCTION COSTS OF NH_3 IN 1978

Feedstock	Assumed cost $/unit	Assumed cost $/MBtu	Unit production cost of NH_3,[a] $/ton
Natural gas	1.00/MBtu	1.00	111
	3.00/MBtu	3.00	174
	5.00/MBtu	5.00	237
Naphtha	0.40/gal	3.00	193
	0.60/gal	4.50	244
	0.80/gal	6.00	294
Fuel oil	0.35/gal	2.43	217
	0.53/gal	3.65	260
	0.70/gal	4.86	302
Coal	20.00/ton	0.95	186
	30.00/ton	1.43	206
	40.00/ton	1.90	225

a. Basis
Plant capacity, 1000 tons/day.
On-stream time, 330 days/yr.

Even with natural gas cost currently ranging up to $2.50/MBtu, it is apparent that alternate feedstocks cannot economically compete with natural gas at the present time. Coal at $20.00/ton ($0.95/MBtu) becomes competitive only after the cost of natural gas exceeds about $3.40/MBtu. It should also be noted that the unit production costs of NH_3 from these feedstocks are based on current levels of technology and the economics of NH_3 production from the various feedstocks in the future will depend not only on the availability and cost of feedstocks, but also on innovations resulting from advanced process development.

POTENTIAL FUTURE SOURCES OF NATURAL GAS

Future supplies of natural gas will be available from either conventional sources (domestic drilling both onshore and offshore) or supplemental sources [imported gas, synthetic natural gas (SNG), or Alaskan gas]. Conventional gas production from known fields is expected to gradually decline in the future as an increasing amount of the gas in these fields is removed. Additional fields will no doubt be discovered but these will typically be in areas which have not been explored and thus where production is expected to be more expensive.

In addition to domestic drilling for gas, numerous other potential sources of natural gas may be used to supplement this conventional natural gas. These sources range from importing gas [Canada, Mexico, and liquefied natural gas (LNG)] to new technologies, such as methane from geopressurized brine, "tight" shales, and coal mines.

Some of these supplemental sources of natural gas are much more likely to be used in the near future than others since the technology is currently available. For example, SNG and imports of natural gas from Canada and Algeria are already being used to provide additional supplies of natural gas (approximately 1.2 Tft^3 in 1978). Other supplemental sources, such as imports from Alaska, Mexico, and Indonesia, may supply additional natural gas in the near future since various contracts are being negotiated at the present time. SNG from coal and other new technologies, although not expected to provide additional gas in the near future, could be contributing significant amounts of gas by the 1990's.

Although technically not a source of natural gas, another mitigating factor in the potential availability of natural gas is the decision of phaseout the use of natural gas in large boilers. About 3 Tft^3 of natural gas (15% of the current supply) is now consumed annually by utility boilers to generate electricity. Although no similar figures are available for consumption by industrial boilers, it is probably at least the same order of magnitude and thus the recent ban on the future use of natural gas as a boiler fuel may free substantial quantities of gas for other uses. In fact, the savings from banning natural gas in utility boilers alone could free twice as much gas as is required to generate all the NH_3 needed domestically in the year 2000.

CONCLUSIONS

A major concern associated with the widespread application of the dry SCR-type NO_x FGT systems in the future is the availability of NH_3. There is both increased emphasis on using coal as boiler fuel and uncertainty surrounding the availability of natural gas, the primary raw material for NH_3 production. As a result, serious questions have been asked about the availability of sufficient quantities of NH_3 for NO_x control without adversely affecting NH_3 prices and thus indirectly impacting the associated agricultural economy.

After reviewing the NH_3 requirements for NO_x control at a typical power plant, the projected U.S. electrical generating capacity, current NH_3 generating techniques, and natural gas availability, the following conclusions have been drawn:

1. NH_3 for NO_x FGT will be purchased on the NH_3 market and not generated at the power plant.

2. Based on the premises of this study, (particularly the assumption that only new large boilers will require NO_x control technology) the widespread application of NO_x FGT systems using dry SCR processes will not have an abrupt adverse impact on the domestic NH_3 market, but rather a gradual increase in demand for NH_3 which can be accommodated by the normal growth of the NH_3 market.

3. If, beginning in 1985, the widespread application of NO_x FGT systems are required on all new large boilers, the annual NH_3 production capacity must be increased from the expected annual growth rate of 3% to 4.5% in order to avoid an increasing (approaching significant) impact on the NH_3 market beyond 1990.

4. At the present time, alternative feedstocks are generally not economically competitive with natural gas as an H_2 source for the production of NH_3.

5. Currently the production of NH_3 consumes only minor amounts of the total U.S. natural gas supply. However, the increased demand for NH_3 due to application of NO_x FGT systems would be contrary to a national energy concern of continuing dependence on natural gas if this NH_3 is based on using natural gas as the feedstock.

INTRODUCTION

Of the five most common air pollutants released in the United States, nitrogen oxides (NO_x) are the only ones projected to increase significantly in the future. The total emission of the other four pollutants [particulate matter, sulfur oxides (SO_x), hydrocarbons, and carbon monoxide (CO)] has decreased during the past 6 yr and is expected to decline further in the future as more point sources are equipped with more efficient control systems. With most public attention focused on these other pollutants which are typically emitted at higher rates than NO_x, the development of equipment and techniques for controlling NO_x has lagged behind. However, recently more attention has been given to the possible health effects of NO_x emissions. Although the NO_x emission from each source appears to be small, the cumulative total is measured in millions of tons annually and is increasing substantially from year to year due to the increasing number of sources.

NO_x emissions are divided into two classes, mobile or stationary, depending on the source. The mobile class, as its name implies, includes sources such as automobiles, trucks, buses, trains, and planes. Attempts at controlling the NO_x emissions from these sources have been delayed due to both technical and economic considerations. Stationary sources, which release about 56% of the total amount of NO_x emitted in the United States (an estimated 11.15 Mtons in 1975), are split into three groups: combustion sources, industrial processes, and other miscellaneous sources. In 1975 combustion sources contributed approximately 93% of all stationary source emission (an estimated 10.4 Mtons) of which nearly two-thirds came from fossil fuel-fired boilers (approximately 7.0 Mtons). If these boilers are further divided according to size (heat rate of the boiler), large boilers [i.e., greater than 250 MBtu/hr (25 MW equiv)] release 45% of al NO_x emitted from stationary combustion sources. Figure 1 and Table 1 show the relative contributions from mobile sources and each type of stationary source to the total NO_x emissions in the United States during 1975. Since the mobile source standards are not expected to become more stringent in the near future, restricting the amount of NO_x emissions from stationary sources, particularly large fossil fuel-fired boilers, may become necessary since these boilers are the second largest source of NO_x emissions in the U.S. The reason for the possibility of stricter NO_x emission control on large boilers are threefold. The emissions from these boilers are projected to increase at about 4.9% annually over the next decade (doubling every 15 yr) assuming present control levels. Secondly, assuming the application of combustion modification which is now considered the best available control technology (about 30% decrease in the NO_x formation), the amount of NO_x emitted by large boilers will increase from 4.7 Mtons in 1975 to 6.6 Mtons in 1985 (50). And thirdly, although the total amount of NO_x emitted by these boilers is only approximately one-half that of mobile sources, there are orders-of-magnitude fewer boilers and hence each boiler source emits significantly more NO_x than each individual mobile source.

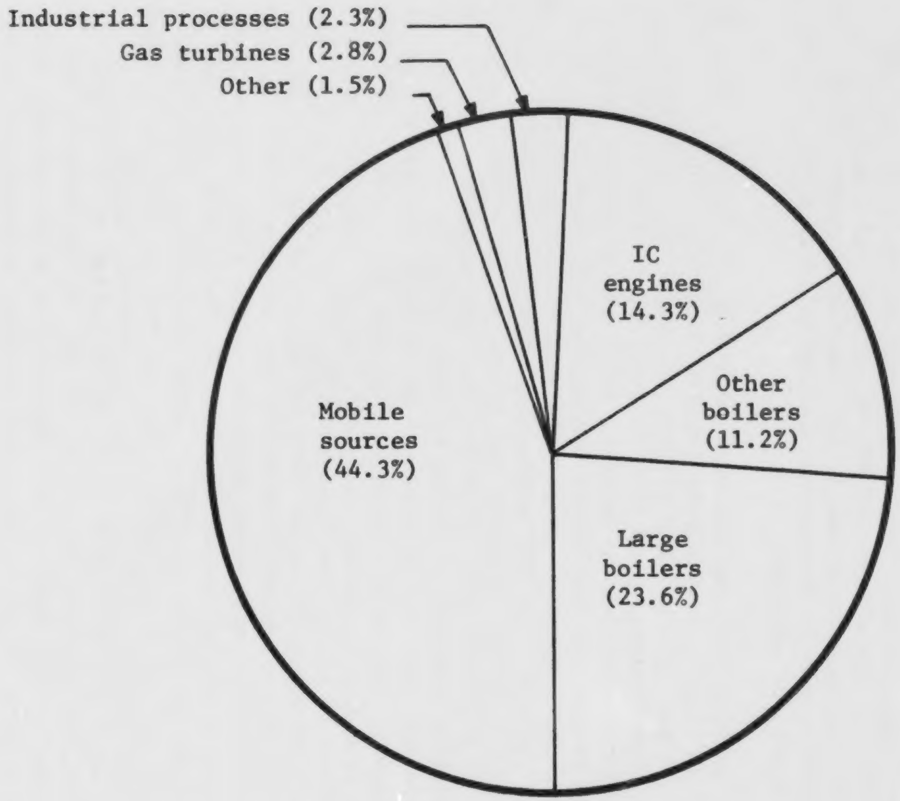

Figure 1. Breakdown of the total U.S. NO_x emissions by original source (45).

Therefore, it is expected that based on each ton of NO_x removed it will be more cost-effective to reduce NO_x emissions from large boilers than from individual mobile sources.

TABLE 1. NO_x EMISSIONS BY SOURCE IN 1975 (45)

	Annual amount, ktons
Mobile sources	8,850
Stationary sources	
Large boilers	4,728
Other boilers	2,238
IC engines	2,849
Gas turbines	566
Incinerators	39
Industrial processes	454
Field burning	290
Total	20,000

FORMATION OF NO_x IN LARGE BOILERS

NO_x is formed during high-temperature combustion operation and can be produced by either of two mechanisms. The first method of formation is thermal fixation or "thermal NO_x" and involves the reaction of molecular nitrogen (N_2) and oxygen (O_2) at high temperatures. The second method is the oxidation of chemically bound nitrogen within the fuel and is called "fuel NO_x." In oil- or coal-fired boilers, NO_x is formed by both mechanisms, but in natural gas-fired boilers the NO_x is formed only by thermal fixation since natural gas contains no fuel-bound nitrogen.

POTENTIAL NO_x CONTROL METHODS FOR LARGE BOILERS

Two potential methods of reducing NO_x emissions from fossil fuel-fired boilers are undergoing extensive research and development. In the United States, combustion modification techniques to reduce the amount of NO_x formed during combustion are being emphasized; while in Japan, mainly because of more stringent NO_x emission regulations, the emphasis is on flue gas treatment (FGT).

Combustion modification techniques attempt to prevent the formation of NO_x by altering the reaction conditions inside the boiler. Since thermal fixation is primarily a function of the temperature (i.e., the rate of formation increases with temperature) and the concentration of O_2 in the boiler, reducing either of these parameters will decrease the amount of thermal NO_x formed. These same boiler modifications which reduce the O_2 concentration in the boiler can also reduce the amount of fuel NO_x formed since the conversion of fuel-bound nitrogen is a strong function of the O_2 concentration. Various combustion modification techniques, which are currently undergoing development work in the United States, are compared in Table 2 for the three types of fossil fuel-fired boilers. Combustion modifications are shown to be the most effective on gas- and oil-fired boilers where a major portion of the NO_x is formed by the thermal mechanism. For coal-fired boilers where much of the NO_x comes from fuel-bound nitrogen, the available combustion modification techniques are limited to about 40% reduction in NO_x.

Although the resulting reduction in the total NO_x emissions from large boilers would appear to be significant, a recent study (45) has determined that, even with combustion modifications on all new large fossil-fueled boilers, the expected growth rate for new large boilers will result in a significant increase in the total NO_x emissions from these boilers, rising from 4.73 Mtons in 1975 to 6.56 Mtons in 1985. Unfortunately, combustion modifications also have undesirable side effects of reducing boiler efficiency, increasing the potential for flame instability, and increasing the soot, CO, and particulate loadings in the flue gas. For these reasons, an alternative method of NO_x control technology, FGT, has recently begun to receive more attention. Although probably more expensive than combustion modification, FGT allows normal operation of the boiler and can remove at least 90-95% of the NO_x from the flue gas. Many different types of FGT systems are currently undergoing development work in Japan and the United States, but the most technically advanced type for removing NO_x from power plant stack gas is now selective catalytic reduction (SCR) (30). In this type, ammonia (NH_3) is injected into the flue gas ducts and the mixture of NH_3 and flue gas passes through a catalytic reactor containing a base-metal catalyst. The NO_x is selectively reduced to molecular N_2 by the following reactions:

$$4NH_{3(g)} + 4NO_{(g)} + O_{2(g)} \rightarrow 4N_{2(g)} + 6H_2O_{(g)} \qquad (1)$$

$$4NH_{3(g)} + 2NO_{2(g)} + O_2 \rightarrow 3N_{2(g)} + 6H_2O_{(g)} \qquad (2)$$

Although these SCR processes are still in early stages of development (most have not been tested either on coal-fired flue gas or on a large scale unit), they are receiving considerable attention because of both the potentially low capital investment and the fact that they generate molecular N_2 directly without further chemical processing.

TABLE 2. EFFECTS OF BOILER MODIFICATIONS TO REDUCE NO_x EMISSIONS BY FOSSIL FUEL TYPE (30)

Boiler modification	Percent decrease of NO_x formation in boilers by fuel type		
	Gas	Oil	Coal
Prevention of thermal NO_x by:			
Flue gas recirculation	60	20	Not effective
Reduced air preheat	50	40	Not competitive
Steam or water injection	60	40	Not competitive
Prevention of both thermal and fuel NO_x by:			
Staged combustion	55	40	40
Low excess air	20	20	20
Reduced heat release rate	20	20	20
Combination of stage combustion, low excess air, and reduced heat release rate	50	35	40
Prevention of fuel NO_x by:			
Change to fuel with lower percent nitrogen	Not applicable	40	20

A primary consideration in the widespread application of these SCR processes in the United States is the potential availability and the cost of NH_3 in the future. Since nearly all of the NH_3 produced in the United States is made from natural gas, there are serious concerns over the potential impact of widespread use of NH_3 in FGT applications. The availability and cost of both natural gas and NH_3 may be affected. Would this additional demand for NH_3 create shortages in NH_3-based fertilizers and thus increase food costs? Will there be sufficient natural gas available for meeting this demand for NH_3? If not, what other methods and feedstocks are available for generating NH_3? This report will attempt to provide insights into these questions.

NH_3 REQUIREMENTS FOR NO_x CONTROL

NH_3 REQUIREMENTS FOR TYPICAL POWER PLANT APPLICATIONS

Any evaluation of the impacts of the additional NH_3 demand for power plant FGT systems must begin with a determination of the magnitude of the additional demand. In order to calculate the NH_3 requirements for typical fossil fuel-fired boilers, it is necessary to presuppose future NO_x standards for large stationary source boilers.

Although the future NO_x emission limits have not been published, for the purposes of this study the following scenario has been hypothesized. Since the SCR and other FGT processes are still in the early stages of development in the United States, strict Federal NO_x emission regulations of 90% removal from large stationary sources will probably not be enforced until 1985. Since Federal new-source performance standards (NSPS) apply only to new sources, existing sources will not be required to meet these standards but existing sources converted from gas or oil to oil or coal after 1985 will be. Between the present and 1985, the NO_x emission limits may be reduced but only to the extent that combustion modifications can be used to meet the newer regulations.

Two possible methods would be available for meeting these emission regulations of 90% NO_x removal, (1) the installation of an SCR system designed for 90% NO_x removal or (2) the use of combustion modifications to cut NO_x emissions from the boiler by 50% with the additional installation of an SCR system designed for 80% NO_x removal to give an overall NO_x removal efficiency of 90%. The impacts of each of these methods are considered in later sections.

The typical fossil fuel-fired boilers included in this study were 500-MW coal- and oil-fired boilers with the flue gas compositions shown in Table 3. Gas-fired boilers were not considered since they are not expected to be built after 1985. For the two alternative cases, the uncontrolled conventional 500-MW coal-fired boiler would release about 600 ppm NO_x in the flue gas (3009 lb/hr), would require an SCR system capable of a 90% NO_x removal efficiency, and would release only approximately 60 ppm NO_x (300 lb/hr) of NO_x after FGT control. Through the use of combustion modification techniques in the second coal-fired case, the boiler would emit 300 ppm NO_x (1504 lb/hr) and the additional SCR system would remove 80% of the remaining NO_x to obtain a total of 90% NO_x removal. In a similar manner, the uncontrolled 500-MW oil-fired boiler releasing 200 ppm NO_x in the flue gas (856 lb/hr) would be required to have 90% NO_x removal, i.e., down to 20 ppm (86 lb/hr) NO_x. Again this could be obtained either by an SCR system designed for 90% removal or through combustion modification to lower boiler emissions to 100 ppm NO_x and then an 80% efficient SCR system.

TABLE 3. FLUE GAS COMPOSITIONS FROM 500-MW

COAL- AND OIL-FIRED BOILERS

Constituent	Flue gas composition from Oil-fired boiler Vol, %	Lb/hr	Coal-fired boiler Vol, %	Lb/hr
N_2	73.60	2,929,000	73.76	3,450,000
O_2	2.54	115,400	4.83	258,200
CO_2	11.96	747,900	12.31	904,200
SO_2	0.13	12,060	0.24	25,130
SO_3	0.0013	151	0.0024	317
NO_x	0.02	856	0.06	3,009
HCl	-	-	0.01	661
H_2O	11.75	300,800	8.79	264,500
	100.00	4,106,000	100.00	4,906,000
Fly ash, gr/sft^3 (wet)		0.032		6.06

The catalytic reduction reactor used in the SCR processes is typically either a fixed-bed reactor incorporating a catalyst or reactor design to prevent plugging by entrained particulates or a moving-bed reactor. The normal operating conditions inside this reactor are a temperature of about 350-400°C (662-752°F) and a space velocity ranging from 5,000-10,000 hr^{-1}. Within these specified conditions the NO_x removal efficiency can be altered by changing the NH_3:NO_x mol ratio as shown in Figure 2. Between 0-90% NO_x removal, the NO_x removal efficiency is nearly a linear function of the NH_3:NO_x mol ratio entering the reduction reactor. Above about 93% NO_x removal the curve becomes asymptotic and greater NO_x removal requires large excesses of NH_3. Thus, for 90% NO_x removal, an NH_3:NO_x mol ratio of about 0.95:1 would be required while for 80% NO_x removal an NH_3:NO_x mol ratio of approximately 0.83:1 would be needed. These mol ratios are based on either laboratory or pilot-plant work of various process developers and most of them recommended operating at a mol ratio in the range of 1.0-1.1:1 for 90% NO_x removal in a commercial system (30).

For this reason the assumed NH_3:NO_x mol ratios for the 90% and the 80% NO_x removal systems were assumed to be 10% higher than that used in the pilot-plant work, i.e., 1.05:1 and 0.91:1 respectively. Once this mol ratio has been estimated, the annual NH_3 consumption for the particular power plant can be obtained from Figure 3. This figure shows the annual NH_3 consumption as a function of the NH_3:NO_x mol ratio for the four hypothesized cases based on the premises given in Table 4: (1) 500-MW coal-fired boiler without combustion modification, (2) 500-MW coal-fired boiler with combustion modification, (3) 500-MW oil-fired boiler without combustion modification, and (4) 500-MW oil-fired boiler with combustion modification. The estimated annual NH_3 consumption for each of these

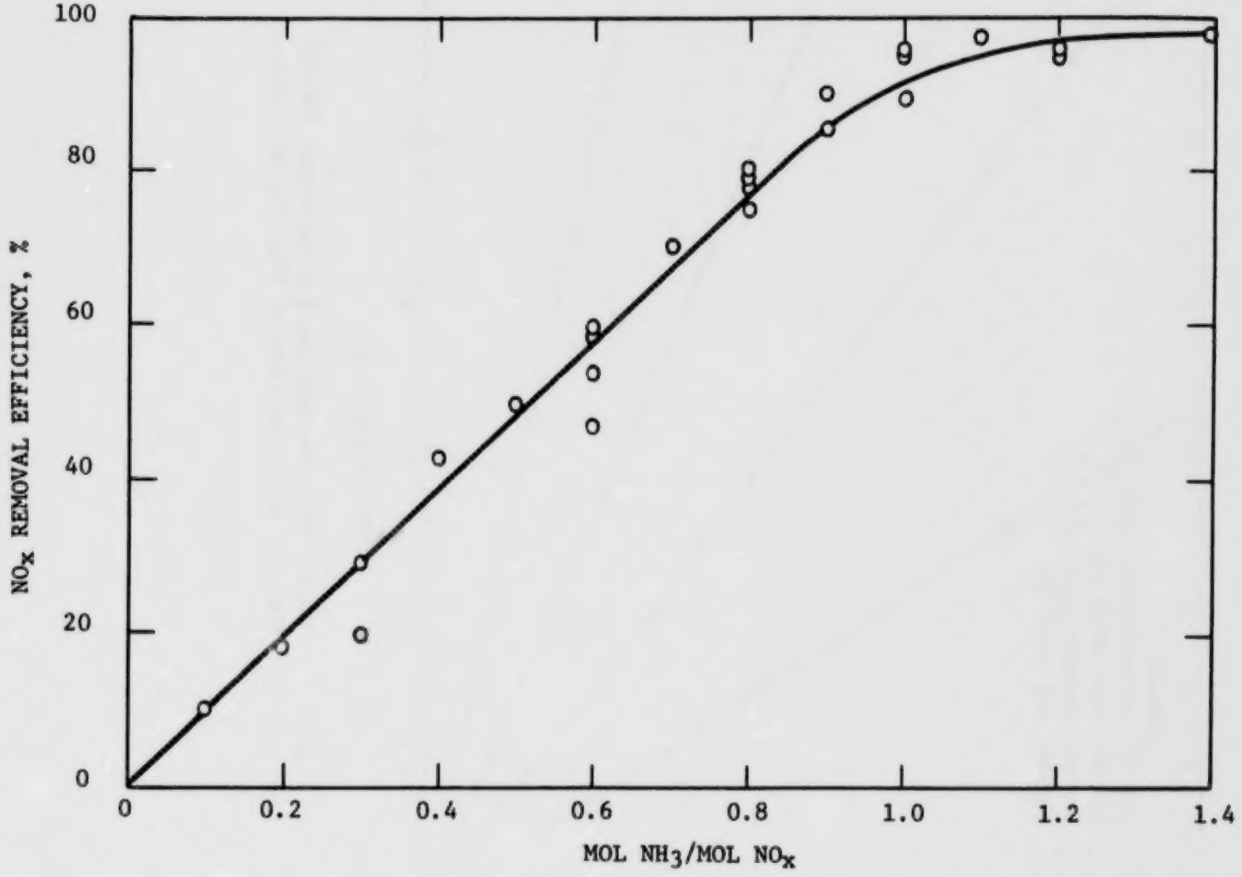

Figure 2. NO_x removal efficiency as a function of NH_3:NO_x mol ratio.

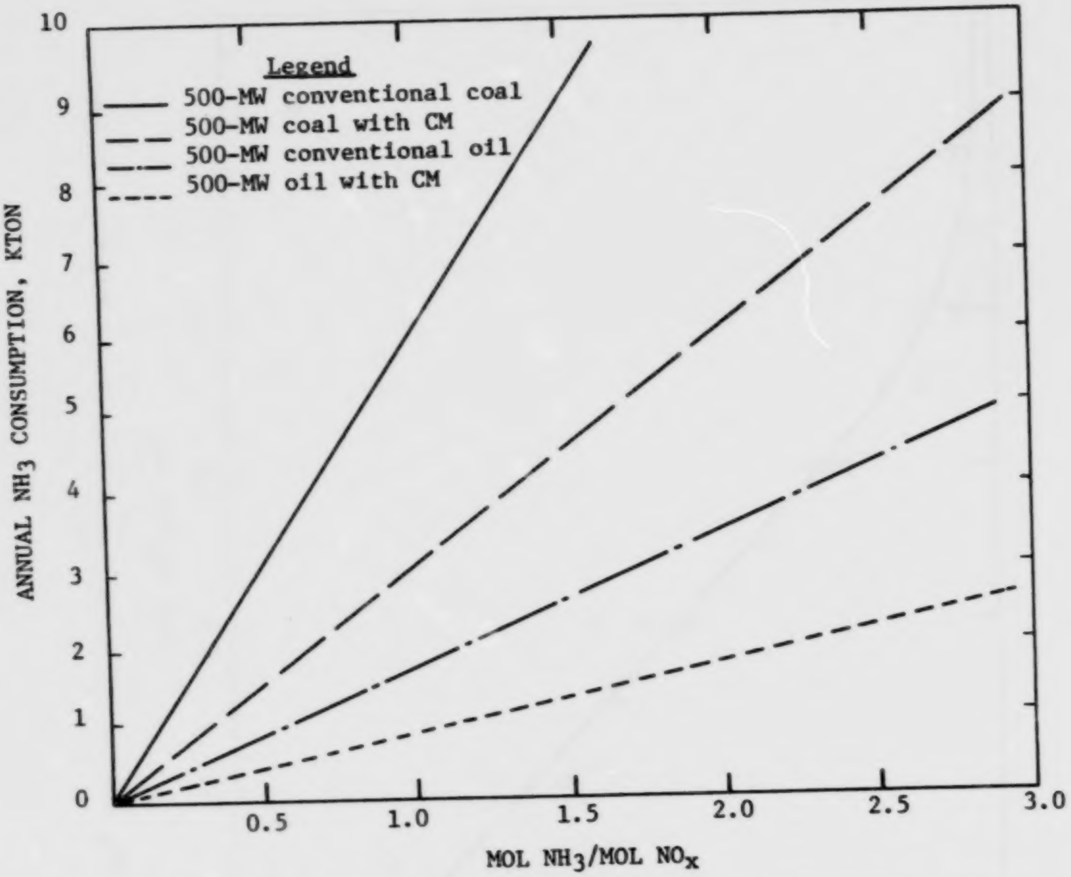

Figure 3. Annual NH_3 consumption for selected 500-MW boilers for NO_x FGT as a function of $NH_3:NO_x$ mol ratio.

four cases is listed in Table 5. The values range from a high of 6266 tons of NH_3/yr for the 500-MW conventional coal-fired boiler without combustion modifications to a low of 772 tons of NH_3 annually for the 500-MW oil-fired boiler with combustion modifications.

TABLE 4. PREMISES FOR THE CALCULATION OF ANNUAL NH_3 CONSUMPTION FOR 500-MW FOSSIL-FIRED BOILERS

Parameter	Value
Heat rate, Btu/kWh	9,000
Heating value	
Coal, Btu/lb	10,500
Oil, Btu/gal	144,000
Availability, hr/hr	7,000

TABLE 5. NH_3 CONSUMPTION FOR TYPICAL 500-MW COAL- AND OIL-FIRED BOILERS

Fuel	NO_x treatment scheme	NO_x concentration in flue gas, ppm	Mols NH_3 per mol NO_x	NH_3 consumption, tons/yr
Coal	SCR[a]	600	1.05	6,266
Coal	CM & SCR[b]	300[c]	0.91	2,715
Oil	SCR	200	1.05	1,781
Oil	CM & SCR	100[c]	0.91	772

a. Selective catalytic reduction.

b. Combustion modification followed by selective catalytic reduction.

c. Combustion modification was assumed to be 50% efficient in controlling NO_x emissions.

The economics of present-day NH_3 generating plants dictate a minimum capacity of 1,000 tons of NH_3/day (330,000 tons/yr) and thus local, small NH_3 plants at each boiler would be highly unlikely. Even with large 2,500-MW coal-fired power plants, the annual NH_3 requirement of 31,000 tons would not justify a captive NH_3 plant. Thus it is assumed that for all NO_x FGT applications the NH_3 would be purchased from existing fertilizer plants already producing NH_3 and shipped to the power plant and stored in large tanks onsite.

PROJECTION OF COAL-FIRED GENERATING UNITS

Any attempt at determining the impact of the widespread application of FGT systems using NH_3 to fossil fuel-fired boilers during the period 1985-2000 requires a reliable projection of new coal- and oil-fired boilers during this period. Various seemingly extraneous factors have a direct bearing on what the total electrical demand will be as well as what proportion of this new electrical capacity will come from which fuel.

In 1977 the U.S. electrical generating industry used the types of fuel shown in Table 6. Numerous projections of changes in this fuel mix have been given by various sources for the period 1975-1985. Although the actual numerical values vary depending on the literature source, several trends are common to all including a drastic increase in nuclear capacity, a moderate increase in coal-fired units, a small increase in oil-fired units, and a reduction in gas-fired units (due to retirements and no new additions after about 1980). These estimates of total installed generating capacity in 1985 range from Energy Research and Development Administration (ERDA) projection of 740 gigawatts (GW) with nuclear power representing 145 GW (1) to the Federal Energy Administration (FEA) 905 GW (the most optimistic of four alternative scenarios provided by FEA) with nuclear supply 150 GW (31). In between these extremes both the Federal Power Commission (FPC) (32) and the National Electric Reliability Council (1) are projecting a 1985 capacity of approximately 798 GW.

TABLE 6. PROJECTED U.S. ELECTRICAL GENERATING CAPACITY

BY FUEL TYPE FOR THE YEARS 1975-1985 (1)

	Generating capacity, GW					
Year	Coal	Oil	Gas	Nuclear	Hydro[a]	Total[b]
1975	183	127	63	33	69	474
1976	193	134	64	43	70	503
1977	204	140	65	51	71	531
1978	218	143	65	59	77	561
1979	232	147	64	69	78	590
1980	247	155	59	78	80	619
1981	260	158	58	93	83	651
1982	278	159	57	113	85	693
1983	291	160	56	133	88	728
1984	308	161	54	153	91	766
1985	321	163	49	174	91	798

a. Includes hydro, pump storage, and other miscellaneous.

b. Columns may not total exactly because of rounding.

These projections of the total generating capacity by fuel type which were made by the National Electric Reliability Council are also shown in Figure 4 for the period 1975-1985. Based on the projections, the net annual capacity additions for each fuel are shown in Table 7. This total U.S. electrical generating capacity of 798 GW and the fuel mix shown in Figure 4 thus provide the basis for projecting the generating capacity during the period 1985-2000.

TABLE 7. ESTIMATED U.S. GENERATING CAPACITY

ADDITIONS BY FUEL TYPE FOR THE PERIOD 1976-1985

	Electrical generating capacity, GW					
Year	Coal	Oil	Gas	Nuclear	Hydro[a]	Total
1975	10	7	1	10	1	503
1976	9	6	1	8	1	531
1977	14	3	0	8	6	561
1978	14	4	(1)[b]	10	1	590
1979	15	8	(5)	9	2	619
1980	13	3	(1)	15	3	651
1981	18	1	(1)	20	2	693
1982	13	1	(1)	20	3	728
1983	17	1	(2)	20	3	766
1984	13	2	(5)	21	0	798

a. Includes hydro, pump storage, and other miscellaneous.

b. Number in parentheses indicates decrease.

Estimates for the period 1986-2000 are more uncertain since both the total generating capacity and the fuel mix are susceptible to economic upheavals or changing trends. The only common trend in the various projections is that the rate of capacity addition is expected to be lower than its historical average of about 6%. Figure 5 projects the U.S. electrical capacity as a function of three growth rates: 4, 5, and 6% with the curve representing 5% probably being the most likely. The portion of this total capacity based on coal as the fuel is expected to increase from the projected 40% in 1985 to about 50% in the year 2000 primarily due to both the long lead times in building nuclear power plants and the elimination of gas and curtailment of oil as boiler fuels. Therefore, for the purposes of this study, the following assumptions have been made:

1. Electrical generating capacity will double during the period 1985-2000. (Average annual growth rate: 4.8%.)

2. Coal-fired power plants will make up 50% of the total generating capacity in the year 2000.

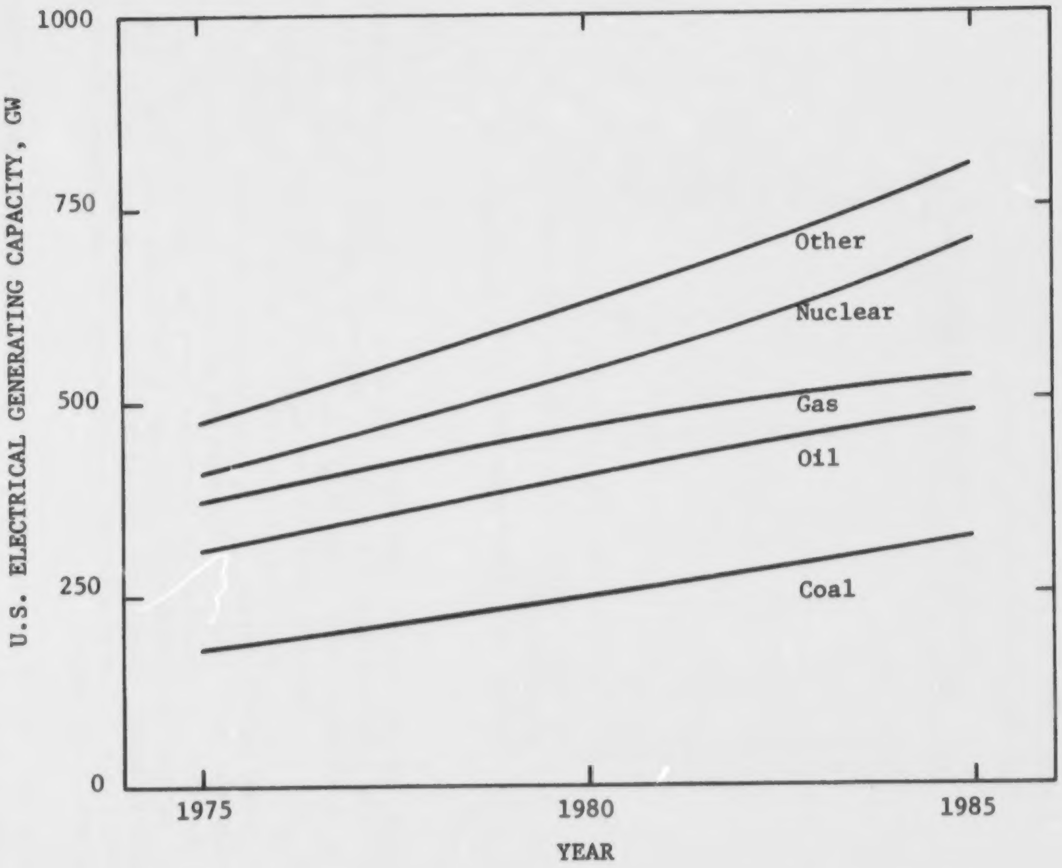

Figure 4. U.S. electrical generating capacity by fuel type for the years 1975-1985 (1).

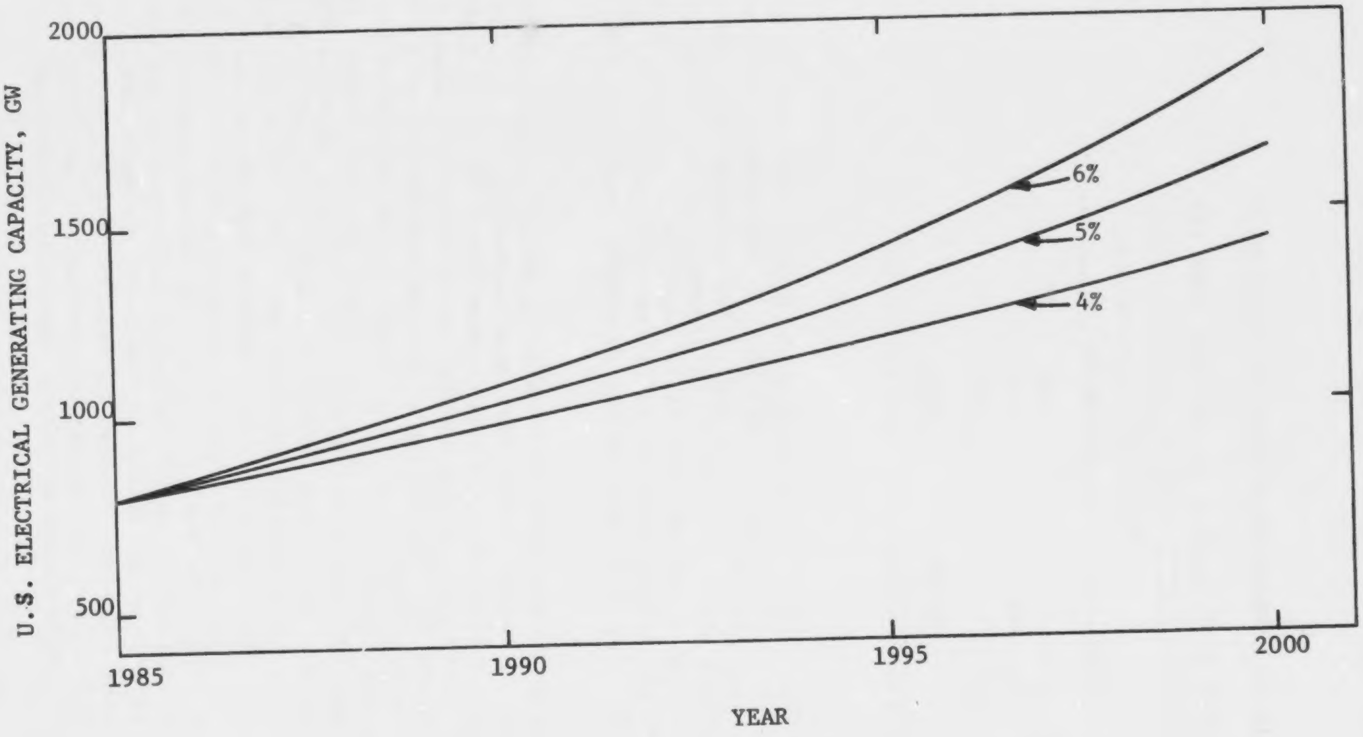

Figure 5. Projected U.S. electrical generating capacity for the period 1985-2000 at three different annual growth rates.

Both of these assumptions are compatible with a report (42) recently released which projects both the total U.S. electrical capacity and also the total coal-fired capacity from 1985 to 2025. Thus for the purposes of this report, the total U.S. generating capacity and the total coal-fired capacity in 2000 were assumed to be 1658 GW and 829 GW as shown in Table 8.

TABLE 8. PROJECTED ELECTRICAL GENERATING CAPACITY BY FUEL TYPE IN THE YEAR 2000 (42)

Fuel	Electrical generating capacity
Coal	829
Nuclear	536
Other	293
Total	1,658

The installed coal-fired capacity during the period 1985-2000 is projected to increase from 321 GW to approximately 829 GW. In addition to this net increase of 508 GW, enough additional new coal-fired capacity will need to be brought online to replace units lost due to retirement. If an annual retirement factor of 0.5% (33) of the installed coal-fired capacity is assumed (i.e., 2-3 GW/yr), the projected coal-fired capacity additions each year for the period 1985-2000 are shown in Table 9 as well as the resulting total installed coal-fired capacity.

These coal-fired units installed and operating after January 1, 1985, will be the plants required to install "best available technology," i.e., 90% NO_x removal. The previously calculated NH_3 requirement for a 500-MW (0.5 GW) coal-fired unit (both case 1 and case 2) was assumed to represent the average NH_3 consumption for new plants and thus multiplying this consumption by the projected new coal-fired capacity requirements given in Table 9, the estimated yearly demand for NH_3 in new coal-fired units was calculated for both cases. These NH_3 requirements for NO_x FGT during the period 1985-2000 are shown in Table 10. For case 1 with all the new coal-fired additions assumed to require 90% SCR, the NH_3 consumption increases gradually at first with the output equivalent to one new 1000 ton/day NH_3 plant being able to supply enough NH_3 to meet all the NO_x FGT applications for the coal-fired units coming online in the United States for the years 1985 and 1986. The number of these NH_3 plants required to produce NH_3 solely for NO_x FGT application would gradually increase from the one plant required in 1985 and 1986 to an estimated 20 plants in the year 2000. For case 2 (all the new coal-fired additions assumed to require only 80% SCR through the use of combustion modifications) the annual NH_3 consumption and hence the number of new NH_3 plants would be about 43% of that required for case 1.

TABLE 9. PROJECTED COAL-FIRED POWER PLANT ADDITIONS AND THE TOTAL COAL-FIRED GENERATING CAPACITY BY YEAR FOR THE PERIOD 1984-2000

Year	Coal-fired capacity additions,[a] GW	Total coal-fired generating capacity, GW
1984	19	308
1985	15	321
1986	23	342
1987	24	364
1988	26	388
1989	27	413
1990	29	440
1991	31	469
1992	33	500
1993	35	532
1994	37	567
1995	40	604
1996	42	644
1997	45	686
1998	48	730
1999	51	778
2000	55	829

a. This average annual increase in coal-fired capacity includes sufficient additional capacity to compensate for an annual retirement rate of 0.5% the total coal-fired generating capacity.

TABLE 10. COMPARISON OF PROJECTED ANNUAL U.S. NH_3 CONSUMPTION BY ALTERNATIVE NO_x FGT SYSTEMS AT ALL NEW COAL-FIRED POWER PLANTS

(Thousands of tons)

	Case 1-Without combustion modifications (600 ppm NO_x-90% removal)		Case 2-With combustion modifications (300 ppm NO_x-80% removal)	
Year	Increase in NH_3 consumption[a]	Total NH_3 consumption for FGT	Increase in NH_3 consumption[a]	Total NH_3 consumption for FGT
1984	0	0	0	0
1985	94	94	41	41
1986	144	332	62	144
1987	150	626	65	271
1988	163	939	71	407
1989	169	1,271	73	551
1990	182	1,622	79	703
1991	194	1,998	84	866
1992	207	2,399	90	1,039
1993	219	2,825	95	1,224
1994	232	3,276	101	1,419
1995	251	3,759	109	1,629
1996	263	4,273	114	1,851
1997	282	4,818	122	2,088
1998	301	5,401	130	2,340
1999	320	6,022	139	2,609
2000	345	6,687	149	2,897

[a] Assuming all new capacity comes online June 1, i.e., only 6 mo of NH_3 required in the first year.

With the recent concern about burning oil and natural gas as boiler fuels, it has been assumed that no new gas-fired utility boilers will be built. Oil-fired boilers, on the other hand, have been used and are expected to continue to be used in areas with significant air quality problems where coal-fired boilers would lead to a marked deterioration of air quality. For this reason new oil-fired generating capacity will be added during the period 1985-2000, but only to a very limited extent. In addition to this small number of new oil-fired sources, new oil-fired boilers would require only about one-third as much NH_3 for NO_x FGT as an equivalent-sized coal-fired unit. Thus when compared with the NH_3 required for NO_x FGT on new coal-fired boilers, the additional NH_3 demand for NO_x FGT on new oil-fired boilers will be negligible and only the NH_3 consumption for coal-fired boilers will be included in this study.

NH_3 Requirements for Both Utility and Industrial Boilers

In formulating the NSPS for NO_x requiring best available technology to be installed on all new large boilers after 1985, i.e., 90% NO_x removal either wholly or partially by NH_3-based SCR treatment, these standards may also be imposed on large industrial boilers (>250 MBtu/hr) as well as utility boilers. In the previous section on NH_3 consumption, only new coal-fired utility boilers were considered primarily because very little information is available on the growth of industrial boilers, particularly when this is further restricted to include only large industrial boilers.

Since so little information is available about these particular boilers, the following general assumption is used in this study: the total NH_3 requirement for NO_x FGT for all large boilers (>250 MBtu/hr) will be 1.25 times the projected NH_3 requirements for NO_x FGT systems on utility boilers. The reasons for selecting the 25% increase are as follows: (1) the total amount of NO_x emitted from all industrial boilers was estimated as 21.6% of that emitted by utility boilers in 1975 and projected to be 22.9% in 1980 (see Table 11), (2) although most of the industrial boilers would have a heat load of less than 250 MBtu/hr, the total amount of NO_x emitted from these smaller boilers would also be proportionately less such that a major portion of the NO_x from industrial boilers would come from the larger boilers, and (3) the impact of these smaller boilers would probably be less since many small boilers would be fired with oil rather than coal.

Table 11 lists the estimated NO_x emissions for 1975 and 1985 from various sizes of boilers (including both utility and industrial boilers) according to the boiler heat load. The large boilers [>250 MBtu/hr (25 MW equiv)] represent more than two-thirds of the total NO_x emissions from stationary source boilers while the midsized boilers (10-250 MBtu/hr) and the small boilers (<10 MBtu/hr), each release about one-sixth of the total NO_x emissions from boilers. Since electrical generating utility boilers were estimated to make up about 80% of these large stationary boilers, multiplying the NH_3 consumption by utility boilers by a factor of 1.25 would provide a reasonable estimate of the total NH_3 demand if 90% NO_x removal by NH_3-based SCR is required for all new large boilers beginning in 1985. With utility boilers making up such a large proportion of these large boilers the growth rate of all large boilers would probably parallel the growth

TABLE 11. ESTIMATED NO_x EMISSIONS FROM VARIOUS SIZED BOILERS IN 1975 AND 1985 (45)

	1975 emissions, Mtons	% of total	1985 emissions, Mtons	% of total	Annual growth rate, %
Large boilers (>250 MBtu/hr)	4.728	67.9	7.605	72.7	4.87
Moderate boilers (10-250 MBtu/hr)	1.297	18.6	1.483	14.2	1.35
Small boilers (0.3-10 MBtu/hr)	0.642	9.2	0.978	9.3	4.30
Boilers (<0.3 MBtu/hr)	0.299	4.3	0.397	3.8	2.88
Total	6.966		10.463		

TABLE 12. COMPARISON OF THE PROJECTED ANNUAL NH_3 DEMAND FOR NO_x FGT DURING THE PERIOD 1985-2000 FOR TWO CASES: REQUIRED ON ONLY NEW UTILITY BOILERS OR FOR ALL NEW LARGE BOILERS

	Annual NH_3 demand, ktons			
	New utility boilers only		All new large boilers	
Year	Case 1[a]	Case 2[b]	Case 1[a]	Case 2[b]
1984	0	0	0	
1985	94	41	118	51
1986	332	144	415	180
1987	626	271	783	339
1988	939	407	1,174	509
1989	1,271	551	1,589	688
1990	1,622	703	2,028	879
1991	1,998	866	2,498	1,082
1992	2,399	1,039	2,999	1,299
1993	2,825	1,224	3,531	1,530
1994	3,276	1,419	4,095	1,774
1995	3,759	1,629	4,699	2,036
1996	4,273	1,851	5,341	2,314
1997	4,818	2,088	6,023	2,610
1998	5,401	2,340	6,751	2,925
1999	6,022	2,609	7,528	3,262
2000	6,687	2,897	8,359	3,622

a. Ninety percent NO_x removal by SCR only.

b. Ninety percent NO_x removal by combined combustion modification and SCR.

rate for large electric utility boilers projected previously and hence overall NH_3 consumption for all large stationary boilers for each year would be 1.25 times that required for only large utility boilers. This total demand for NH_3 for NO_x FGT on all large boilers is shown in Table 12.

CURRENT NH_3 SUPPLY AND DEMAND

The earliest and still the most important use of NH_3 is for the production of fertilizers. Approximately 75% of the current U.S. production of 16.4 Mtons of NH_3 is used either directly or indirectly as a fertilizer with the remaining 25% apportioned as follows: 10% for fiber and plastic intermediates, 10% for commercial explosives, and the remaining 5% for exports (9).

Over the past 10 yr the NH_3 demand in the United States, as shown in Table 13, has steadily increased at an annual rate of about 4.5% from 10.6 Mtons in 1966 to 16.4 Mtons in 1976. However, the combination of a very low annual growth rate in the agricultural use of NH_3 (2-3%), a low demand for fiber and plastic intermediates, and the projected absence of export demands have led others to project an annual growth rate of only 3% for NH_3 during the period 1976-2000 (29). The use of NH_3 for commercial explosives has the potential for large increases in the future due to the expected increase in coal mining but since this represents such a small portion of the NH_3 market its overall impact on NH_3 consumption is expected to be negligible.

TABLE 13. U.S. NH_3 DEMAND 1965-76 (4, 39)

Year	NH_3 demand, ktons Fertilizer[a]	 Total[b]
1965	5,633	8,869
1966	6,468	10,605
1967	7,319	12,194
1968	8,242	12,120
1969	8,449	12,769
1970	9,057	13,824
1971	9,877	14,538
1972	9,741	15,169
1973	10,070	15,208
1974	11,120	15,733
1975	10,450	16,393
1976	12,560	16,423

a. Crop year.
b. Calendar year.

For comparison purposes, Figure 6 shows the NH_3 demand for the period 1976-2000 under three alternative annual growth rates: 3, 4, and 5%. The projected demand (excluding the NH_3 demand for NO_x FGT) in the year 2000 ranges from 34.3 to 55.5 Mtons depending on the growth rate assumed. Although the historical trend has averaged about 4.5%, the projected NH_3 demand is expected to increase at a lower growth rate of 3.0%. Under this assumption the NH_3 demand in 1985 and 2000 would be 21.4 and 33.1 Mtons respectively.

The U.S. NH_3 production capacity on the other hand has grown primarily by boom-bust cycles in recent years. For example, up until the mid-1960's NH_3 plants were typically small plants (<300 tons/day) located near the ultimate consumers. However, with the advent of large capacity centrifugal compressors, most new NH_3 plants jumped in size to the 1000 ton/day plant which is common today. These plants offered distinct economic advantages over the small plants and rapidly drove most of these smaller plants out of business when the resulting surplus capacity came online. This overcapacity and subsequent depressed market price for NH_3 also resulted in very little incentive to add further capacity during the early 1970's. However, the constantly increasing demand for NH_3 coupled with the beginning of natural gas curtailment during the winter months rapidly absorbed the overcapacity until in 1974 and 1975 the United States became a net importer of NH_3 and the wholesale price of NH_3 doubled and even tripled within 1 yr (54). With these economic incentives numerous NH_3 producers announced plant expansions and additions which have come online during 1977 and 1978. This large increase in capacity had led to another oversupply situation in the domestic NH_3 market. Table 14 lists the NH_3 capacity additions and the U.S. NH_3 plant capacity by year during the period 1967-1980. The operating rate of the U.S. NH_3 plants during this same period is shown in Figure 7. As would be expected during periods when new NH_3 capacity is coming online, the operating rate of the NH_3 industry declines until the demand soaks up excess capacity.

In view of the past history of fluctuating capacity growth, projecting NH_3 plant capacities expected for the year 1980-2000 are rather difficult. Various factors which affect decisions to increase capacity include (1) government policy and (2) current market conditions, which can change rapidly. Due to the sensitivity of the NH_3 market to supply and demand, the NH_3 plant capacity is expected to grow intermittently whenever demand, and hence prices, rise and provide economic incentives for expansion. A reasonable assumption in the projection of growth is that the total U.S. capacity will keep pace with the projected increase in demand, i.e., approximately 3%/yr. Figure 8 shows the capacities for the period 1980-2000 based on a growth rate of 3.0% as well as 4% and 5% growth for comparison purposes. The projected NH_3 capacity in 1985 and 2000 based on a 3% annual growth rate would be 27.9 and 43.5 Mtons respectively.

COMPARISON OF PROJECTED NH_3 DEMAND FOR NO_x FGT WITH POTENTIAL NH_3 SUPPLY

The previously calculated NH_3 demand for NO_x FGT (90% NO_x removal) required beginning in 1985 for all new large boilers (industrial as well as utility boilers) is compared with the projected NH_3 demand for the period 1985-2000 in Table 15 and the relative NH_3 demands for NO_x FGT and conventional uses as shown in Figure 9.

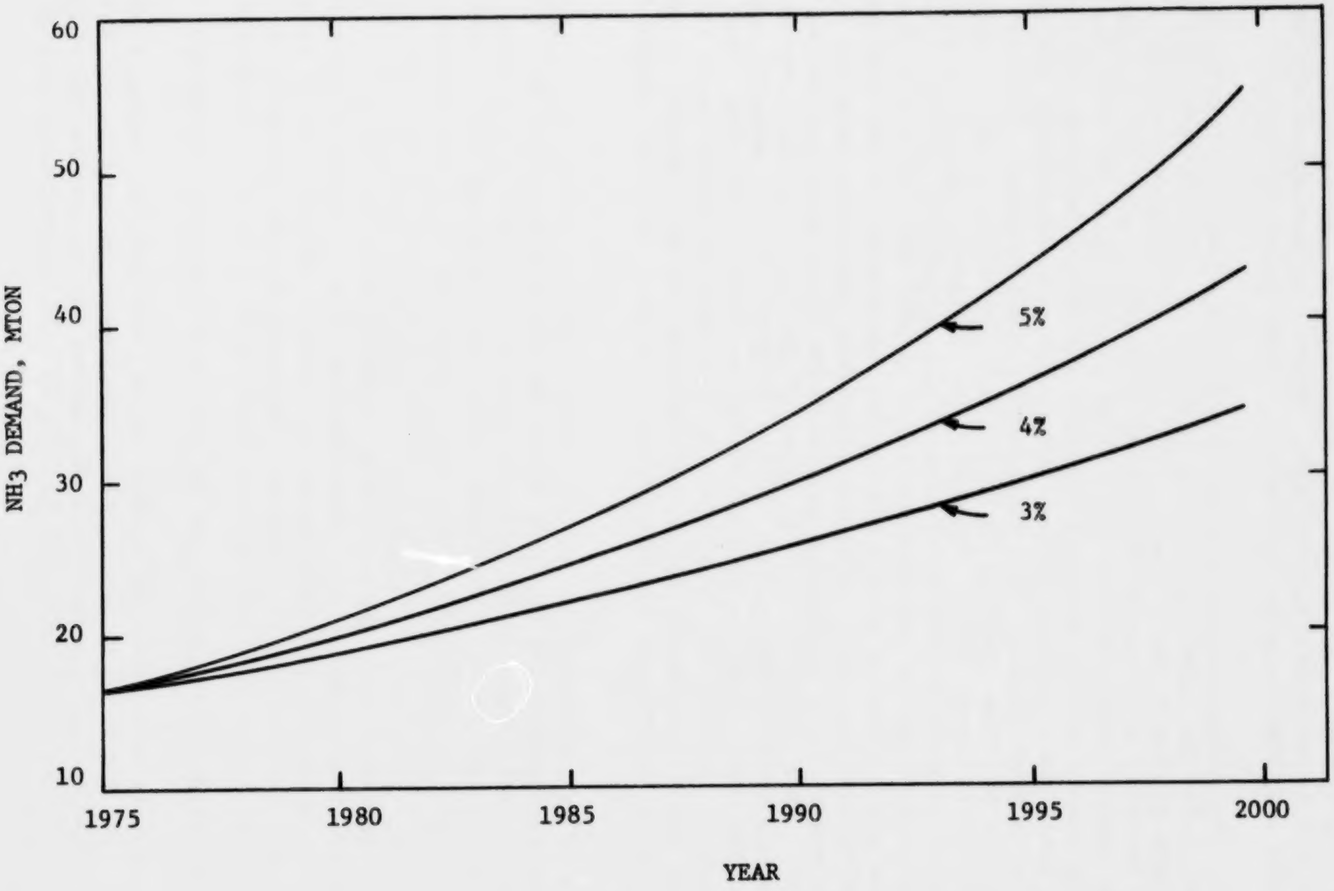

Figure 6. Projected U.S. NH_3 demand during the period 1975-2000 at three annual growth rates.

TABLE 14. U.S. NH_3 PLANT CAPACITY ADDITIONS AND TOTAL U.S. NH_3 PLANT CAPACITY DURING THE PERIOD 1967-1980 (26)

Year	Additional capacity, ktons[a]	Total capacity ktons
1967	-	12,991
1968	1,921	14,806
1969	103	14,848
1970	484	16,332
1971	(30)	16,302
1972	45	16,347
1973	645	16,992
1974	(38)	16,954
1975	1,203	18,157
1976	455	18,703
1977	2,701	21,404
1978	1,337	23,156
1979	951	24,107
1980	0	24,107

a. Parentheses indicate decrease.

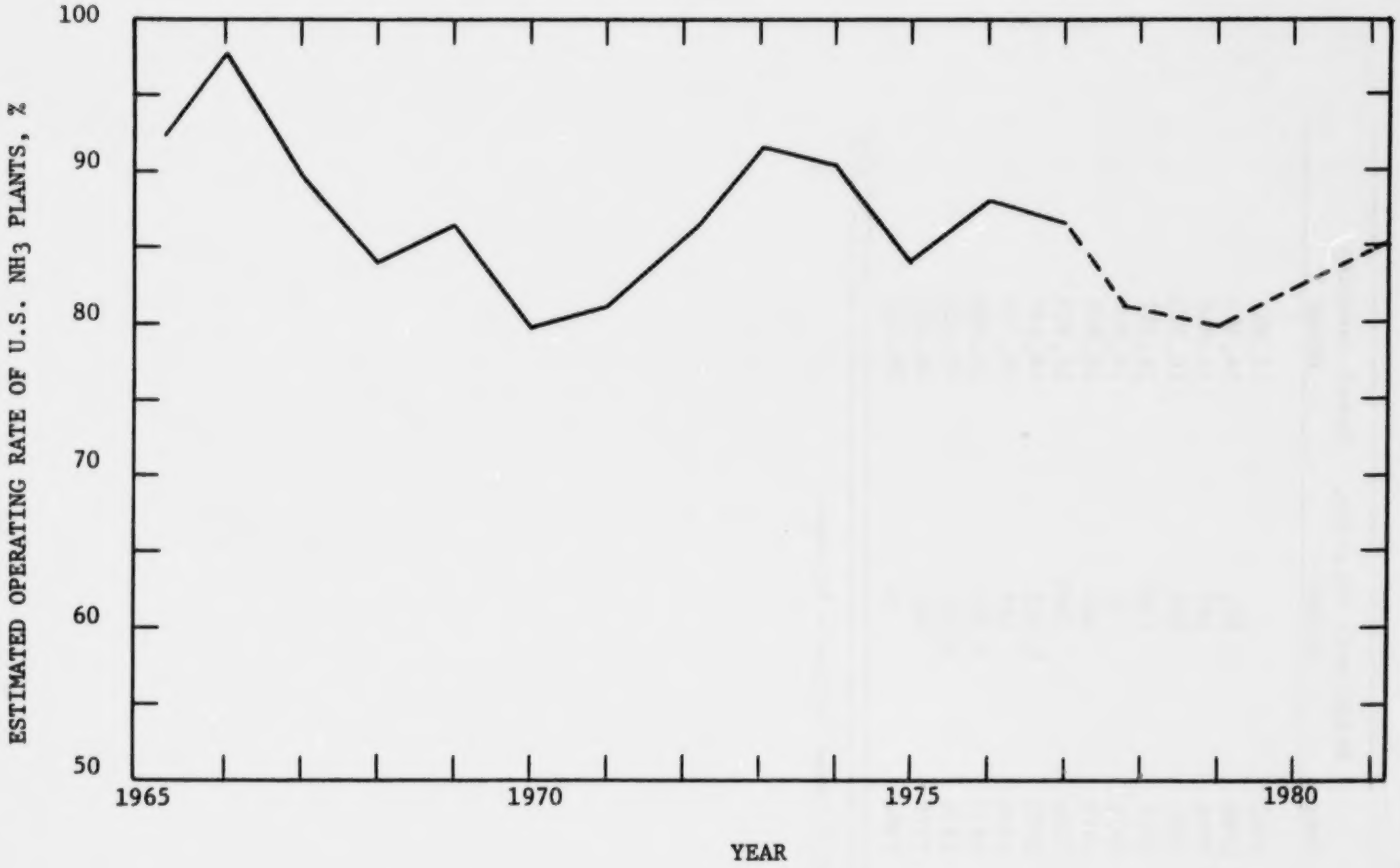

Figure 7. Annual operating rate of U.S. NH_3 plants from 1965-1981 (43).

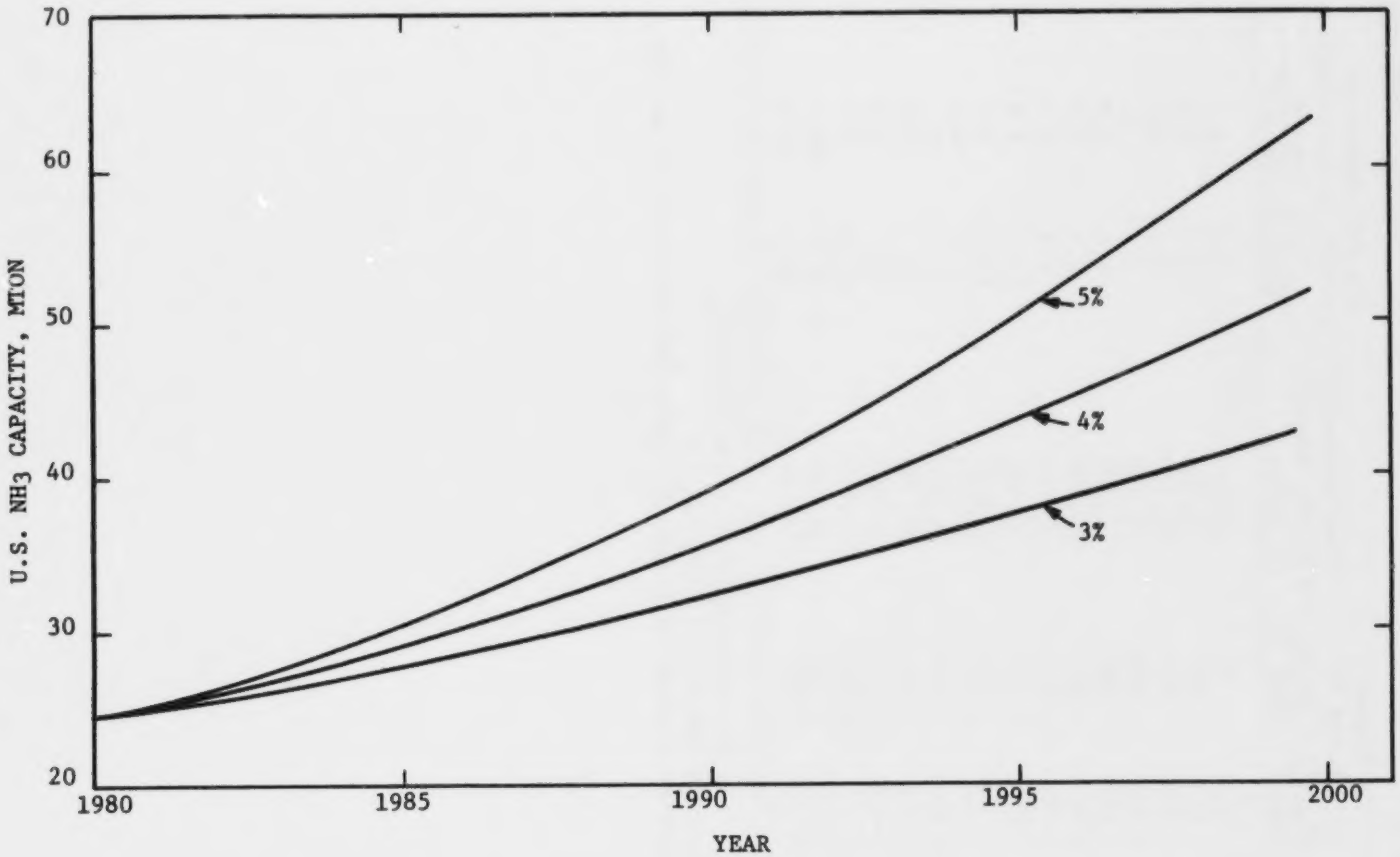

Figure 8. Projected U.S. NH3 capacity during the period 1975-2000 at three annual growth rates.

TABLE 15. COMPARISON OF THE ESTIMATED ANNUAL NH_3 DEMAND FOR NO_x FGT WITH THE PROJECTED CONVENTIONAL U.S. NH_3 DEMAND DURING THE PERIOD 1985-2000

Year	Estimated NH_3 demand for NO_x FGT,[a] ktons		Total projected U.S. NH_3 demand, ktons	NO_x FGT demand as percent of projected U.S. demand	
	Case 1[b]	Case 2[c]		Case 1[b]	Case 2[c]
1985	118	51	21,400	0.6	0.2
1986	415	180	22,100	1.9	0.8
1987	783	339	22,700	3.4	1.5
1988	1,174	509	23,400	5.0	2.2
1989	1,589	688	24,100	6.6	2.9
1990	2,028	879	24,800	8.2	3.5
1991	2,498	1,082	25,600	9.8	4.2
1992	2,999	1,299	26,400	11.4	4.9
1993	3,531	1,530	27,100	13.1	5.6
1994	4,095	1,774	28,000	14.8	6.3
1995	4,600	2,036	28,800	16.5	7.1
1996	5,341	2,314	29,700	18.2	7.8
1997	6,023	2,610	30,600	19.9	8.5
1998	6,751	2,925	31,500	21.6	9.3
1999	7,528	3,262	32,400	23.2	10.1
2000	8,359	3,622	33,400	25.3	10.8

a. Coverage: all new large boilers (>250 MBtu/hr).
b. Ninety percent removal by SCR only.
c. Ninety percent removal by combined combustion modification and SCR.

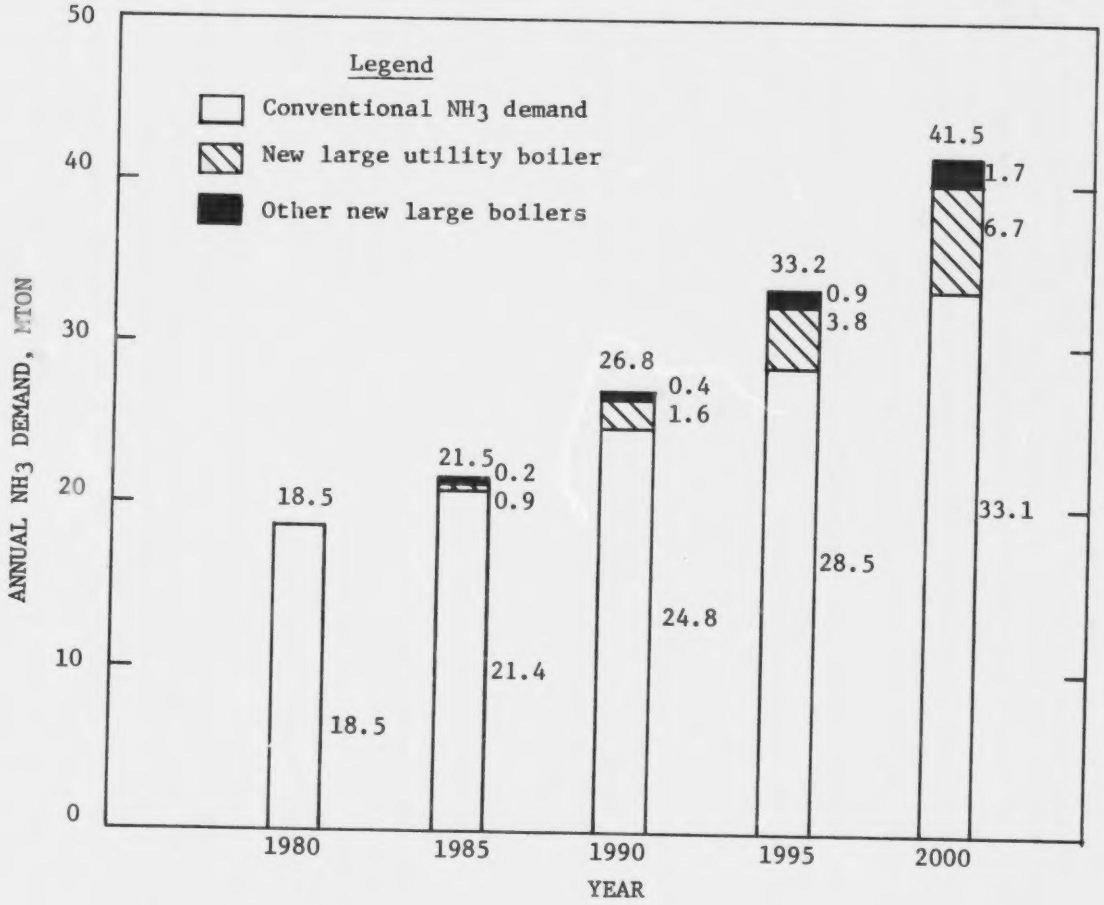

Figure 9. Comparison of the projected NH_3 demand for NO_x FGT with conventional NH_3 demand for selected years during the period 1980-2000 (case 1 only).

The additional NH_3 demand for NO_x FGT for case 1 (90% NO_x removal by SCR only) initially would require only 0.6% of the projected NH_3 supply in 1985 but would steadily increase until it requires slightly more than 25% of the total supply in 2000. For case 2 (90% NO_x removal by combined combustion modification and SCR) these percentages range from 0.2% in 1985 to nearly 11% in the year 2000. In either case these increases are slow enough that with enough foresight, represented by the 1.5-2.0 yr lead time required to build these plants, the NH_3 plant capacity could keep pace with the increasing NH_3 consumption.

Thus under the assumptions used in this study, particularly the assumption that only new sources operating after 1985 are required to obtain 90% NO_x removal through the use of FGT, the application of the best available technology rule for controlling NO_x emissions from large boilers will not result in an abrupt disruption of the NH_3 market in which the demand for NH_3 suddenly increases above the available supply and thus has an adverse impact on the availability and price of NH_3. In fact, with the proper amount of initial planning the potentially disruptive effects of the application of NO_x FGT on the NH_3 industry can be minimized. Under these assumptions the primary impact on the domestic NH_3 market would be to cause the U.S. NH_3 demand to increase at 4.5%/yr during the period 1985-2000 rather than the previously assumed 3.0%.

However, various alternative assumptions could result in substantially different scenarios. For example, continuing delays in nuclear plant construction could result in an increased emphasis on coal-fired power plants during the period rather than the 50-30 mix of coal and nuclear. For example, a 5% reduction in total nuclear capacity (about 27 MW) would probably result in 3.3% (or 27 MW) increase in both the total coal capacity and the annual NH_3 consumption. Another critical assumption is a doubling of the total U.S. generating capacity between 1985-2000, i.e., an annual increase in capacity of about 4.7% rather than a higher historic level of 6%. A significant portion of the additional electrical capacity under a 6% growth rate would be based on coal and thus there would be a higher than projected NH_3 demand.

A more significant assumption is that only new coal-fired power plants coming online after 1985 are required to use NH_3-based NO_x FGT systems. If, as shown in Table 16, all coal-fired power plants were required to install a 90% efficient NH_3-based NO_x FGT system, NH_3 consumption would increase by about 4 Mtons in 1985 (from approximately 1 new 1000 ton/day plant to 11 plants) and 10.4 Mtons in 2000 (from about 20 new plants to 32 plants). This immediate and large increase in NH_3 demand, equivalent to approximately 19% of the projected NH_3 demand in 1985, could create significant economic problems in the agriculture industry.

The availability of NH_3 for the widespread application of NO_x FGT systems to new power plants after 1985 does not seem to be a potentially serious problem since this additional demand is, at least in the early stages, relatively small compared with conventional NH_3 demands for fertilizer applications. The primary requirement in minimizing the impact on the NH_3 market will be to gradually increase domestic NH_3 supplies at a rate of 4.5%/yr rather than the currently projected 3%. Since most of the current domestic supply of NH_3 is generated from natural gas, the answer to the question of the future availability and cost of natural gas will have a major impact on the availability and cost of NH_3.

TABLE 16. PROJECTED U.S. NH_3 DEMAND FOR NO_x FGT FOR THE PERIOD 1985-2000 IF ALL COAL-FIRED POWER PLANTS REQUIRED TREATMENT

Year	Total coal-fired generating capacity, GW	NH_3 consumption for NO_x FGT,[a] Mtons	Percent of projected total U.S NH_3 production
1985	321	4.02	18.8
1986	342	4.29	19.4
1987	364	4.56	20.1
1988	388	4.85	20.8
1989	413	5.18	21.5
1990	440	5.51	22.2
1991	469	5.88	23.0
1992	500	6.27	23.7
1993	532	6.67	24.8
1994	567	7.11	25.7
1995	604	7.57	26.6
1996	644	8.07	27.4
1997	686	8.60	28.4
1998	730	9.15	29.3
1999	778	9.75	30.4
2000	829	10.39	31.4

[a] Basis
FGT system, 90% NO_x removal by SCR.

Although the total amount of natural gas consumed in producing NH_3 does not represent a significant portion of the current or projected natural gas supply, under the present conditions with gas curtailments during the winter months and the situation projected to deteriorate in the future, it is uncertain whether additional amounts of natural gas will be available for generating NH_3 in the future. For example, in 1978 total domestic NH_3 production is estimated at 17.4 Mtons which will consume approximately 0.54 trillion cubic feet (Tft^3) of natural gas, about 2.7% of the total U.S. gas production of 20 Tft^3.

In the year 2000 the production of NH_3, even without the additional demand for NO_x FGT systems, will more than double and hence could require significant portions of the total U.S. natural gas supply. Serious questions have been raised as to whether this amount of natural gas will be available for NH_3 production. Thus the major impact of the widespread application of NO_x FGT systems may not be the direct impacts on the NH_3 market but rather the indirect impacts on the availability and price of natural gas. The following sections will discuss the current hydrogen (H_2) feedstocks for NH_3 and projections of the future availability of these current sources and other potential H_2 feedstocks.

CURRENT NH_3 GENERATING TECHNIQUES

STEAM REFORMING OF NATURAL GAS (27, 53)

Although NH_3 has been made by other methods in the past, the simplest procedure is to chemically combine molecular N_2 and molecular H_2. This N_2 can be obtained by simply filtering and injecting air into the system. Process alternatives to NH_3 production actually revolve around the method in which H_2 is produced. The H_2 must either be purchased directly, such as from a chlorine (Cl_2) manufacturer, or more commonly, generated by chemical reaction. This latter method is causing the most interest since the most common chemical method is the steam reforming of natural gas. In fact about 95% of the NH_3 currently being produced in the United States is based on this method (47). Since steam reforming of natural gas represents such an overwhelming majority of the NH_3 plant capacity, a state-of-the-art review of current NH_3 generating techniques will be primarily concerned with this method.

Natural gas became the primary feedstock during the 1950's with the rapid development of large commercial gas fields in Texas, Louisiana, and Oklahoma and the completion of the transcontinental pipelines. Natural gas, primarily containing methane as shown in Table 17, is an excellent source of H_2 since it is a relatively clean hydrocarbon feedstock and also has a high ratio of hydrogen to carbon. These two factors combined meant that a high-quality feedstock was readily available through the United States and other NH_3 feedstocks such as coal were rapidly replaced by natural gas.

TABLE 17. TYPICAL NATURAL GAS COMPOSITION (53)

	% by vol	
Constituent	At wellhead	As delivered
CH_4	74.9	92.3
N_2	-	2.0
Ar	-	0.4
CO_2	7.3	0.01
H_2S	8.9	<5 ppm
Hydrocarbons		
C_2	3.2	3.2
C_3	1.1	0.9
C_4	0.8	0.2
C_5	0.5	0.01
C_6 or higher	2.3	-

Processing Scheme

A general outline of the processing scheme in a typical natural gas-based NH_3 plant is shown in Figure 10. Although the natural gas from the pipeline is usually of high purity, the gas may initially be passed through a gas purification system where any remaining sulfur compounds, primarily hydrogen sulfide (H_2S) are removed by adsorption. Activated carbon is typically used because activated carbon operates at ambient temperatures and can be regenerated. When the sulfur content is unusually high, another method such as hot zinc oxide (ZnO) must be used. Even though the sulfur content is already low when pipeline gas is used, this purification step is included to prevent the poisoning of the nickel (Ni) catalyst used in the steam reforming of the natural gas.

The purified natural gas is injected with steam and passed through a two-stage reformer system, a primary reformer (a sketch of which is shown in Figure 11), and a secondary reformer with both containing a bed of Ni catalyst. Most of the methane is converted to H_2 and carbon dioxide (CO_2) by the following reactions in the primary reformer.

$$CH_{4(g)} + 2H_2O_{(g)} \rightleftarrows 4H_{2(g)} + CO_{2(g)} \quad (1)$$

$$CH_{4(g)} + H_2O_{(g)} \rightleftarrows 3H_{2(g)} + CO_{(g)} \quad (2)$$

$$CO_{(g)} + H_2O_{(g)} \rightleftarrows CO_{2(g)} + H_{2(g)} \quad (3)$$

The operating conditions are about 25-35 atmospheres pressure, temperatures of 780-850°C (1436-1562°F), and a steam to carbon mol ratio of 3.5-4.0:1. These high temperatures in the reformer are maintained by indirect heating using both the purge gas from the synthesis loop and supplemented natural gas as fuel. The catalyst in these tubes is Ni on a fused alumina (Al_2O_3) base with the Ni content ranging from 12 to 22%. Because of the extreme operating conditions in the primary reformer, unusual materials of construction for the catalyst tube are required such as specialty stainless steel.

The gas from the primary reformer, containing some residual methane, is injected with air and sent to a secondary reformer. The amount of air introduced into the system is controlled such that the final H_2:N_2 mol ratio will be 3:1 when the makeup gas enters the NH_3 synthesis loop. The Ni catalyst [typically 14% Ni, 15% calcium oxide (CaO), and 67% Al_2O_3] in the secondary reformer operates in the temperature range of 1150-1204°C (2100-2200°F). A typical gas composition entering and leaving the secondary reformer is given in Table 18.

CO Conversion

CO present in the reformed gas is a potential source of H_2 and a poison to the NH_3 synthesis catalyst currently in use and necessitates the inclusion of a conversion unit. The conversion section is a two-stage unit consisting

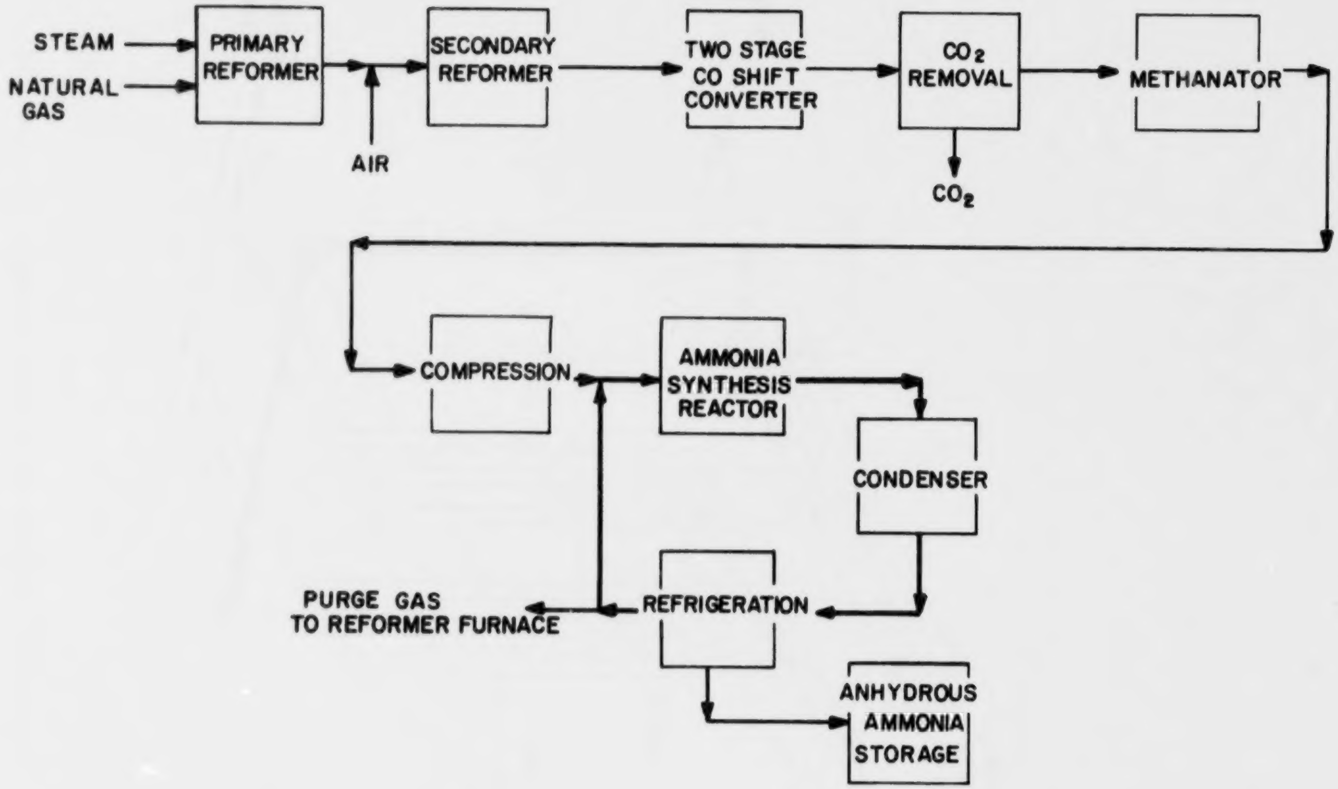

Figure 10. Block flow diagram of a typical NH_3 plant based on natural gas as the H_2 feedstock.

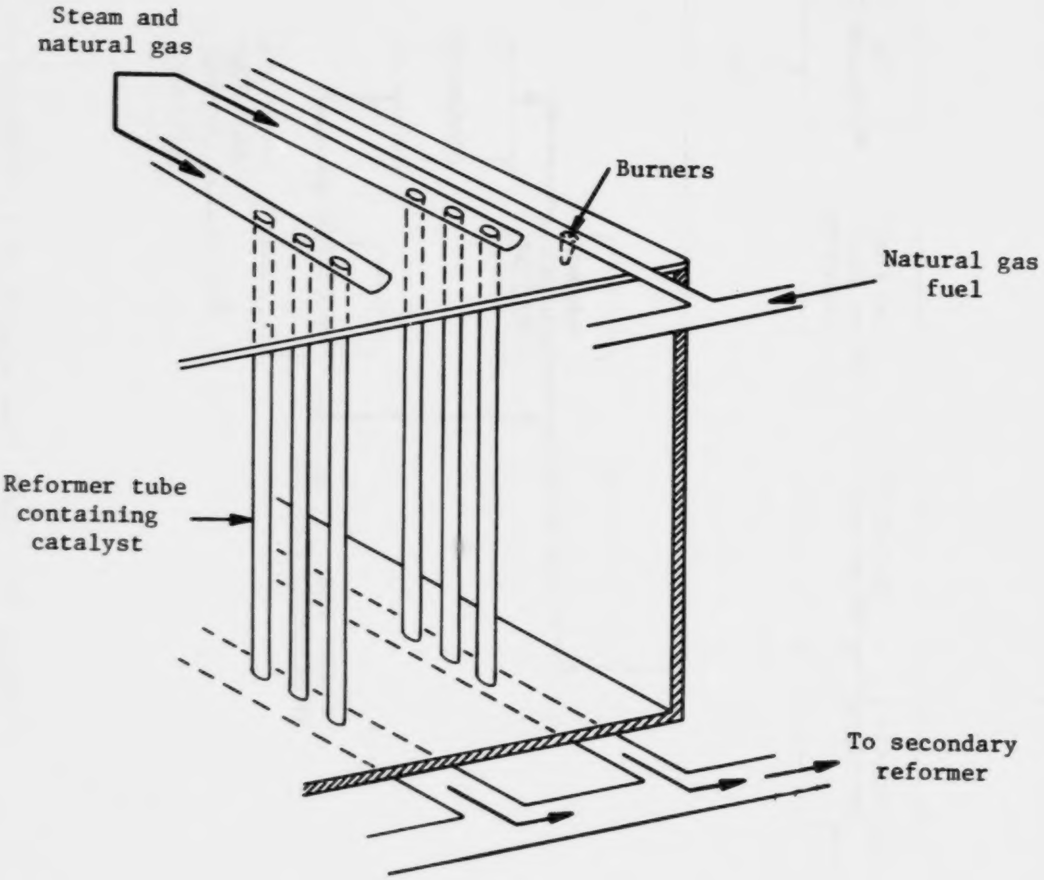

Figure 11. Diagram of typical primary reformer (53).

of a high temperature and low temperature converter with an interstage heat exchanger. Although a single-stage unit could be used to convert the CO back to methane, this would lead to a high consumption of H_2 since 3 mols of H_2 are required to convert 1 mol of CO to methane.

$$CO_{(g)} + 3H_{2(g)} \rightleftarrows CH_{4(g)} + H_2O_{(g)} \qquad (4)$$

The reduction of CO to methane also negates the potential water gas shift reaction of CO to produce additional H_2, as shown in reaction (3). For these reasons, the first-shift stage is a high-temperature reactor using an iron-chromium oxide (Fe_2O_3-CrO) catalyst (55% Fe and 6% Cr) at 415°C (750°F) to promote the water gas shift (reaction 3) and thereby consuming as much CO as possible and at the same time maximizing the production of H_2.

TABLE 18. TYPICAL GAS COMPOSITION ENTERING AND LEAVING SECONDARY REFORMER (53)

Constituent	% by vol, dry basis Entering	Leaving
H_2	71.4	58.1
CO	5.5	8.5
CO_2	14.4	11.5
CH_4	8.7	0.2
N_2	-	21.3
Ar	0.03	0.3
O_2	-	0.1

After cooling to 250°C (482°F) in a heat exchanger the gas is then passed through the low-temperature catalytic reactor containing zinc (Zn), copper (Cu), and sometimes Al_2O_3 to further reduce the CO exit concentration by reaction (3) to approximately 0.3%. In addition to the interstage heat exchange an absorbent cartridge of ZnO is usually used before the second-shift stage to remove the last traces of any sulfur- or chlorine (Cl)-containing compounds since the low-temperature catalyst is extremely susceptible to sulfur and Cl poisoning.

Once most of the CO has been shifted to CO_2, the CO_2 is removed from the systems, usually by solvent extraction. Although monoethanolamine (MEA) was used in earlier low-pressure NH_3 plants, a "promoted" hot potassium carbonate (K_2CO_3) method such as the Vetrocoke, Benfield, or Catacarb processes have come into increasing use with the development of modern higher pressure NH_3 plants. These newer carbonate processes are labeled as "promoted" since an additive has been added to the carbonate solution which increases the rate of CO_2 absorption.

The synthesis gas entering the CO_2 removal system is relatively hot (90-100°C) and passes countercurrently to a hot K_2CO_3 solution in a two-stage absorber. The CO_2 is absorbed and reacts with the K_2CO_3, i.e.,

$$K_2CO_{3(aq)} + CO_{2(g)} + H_2O_{(g)} \rightleftarrows 2KHCO_{3(aq)} \qquad (5)$$

The exiting gas from the second stage of the CO_2 absorber typically contains 0.02-0.07% CO_2.

The CO_2-rich carbonate solution from the absorber is partially regenerated by releasing the pressure in a power recovery turbine. The solution is then pumped to a two-stage regeneration vessel. The upper section of the vessel is simply a flashing tank where most of the CO_2 separates from the solution and passes overhead. A portion of the resulting lean solution recycles to the first stage of the absorber while the remainder drops to a second stage of the regenerator. The second stage uses indirect steam heating to further strip the carbonate solution of CO_2 prior to recycle to the second stage of the absorber.

Final Purification

Due to the sensitivity of the NH_3 synthesis catalyst to oxides of carbon, the remaining CO, CO_2, and O_2 concentrations must be further reduced to low levels (typically <10 ppm). This final gas purification step is usually an exothermic catalytic methanation reaction for CO and CO_2 and a hydrogenation reaction for oygen, i.e.,

$$CO_{(g)} + 3H_{2(g)} \rightarrow CH_{4(g)} + H_2O_{(g)} \qquad (7)$$

$$CO_{2(g)} + 4H_{2(g)} \rightarrow CH_{4(g)} + 2H_2O_{(g)} \qquad (8)$$

$$2H_{2(g)} + O_{2(g)} \rightarrow 2H_2O_{(g)} \qquad (9)$$

These reactions occur at about 230-315°C (446-600°F) over a nickel oxide (NiO) catalyst; hence heat exchange between the inlet and outlet gases is needed to preheat the inlet gas up to the reaction temperature. The typical inlet and outlet gas compositions around the methanator are shown in Table 19.

The resulting purified synthesis gas is sent to centrifugal compressors which boost the gas to the NH_3 loop pressure of 2000-4000 psig (137-273 atm). This increase in pressure is usually carried out in three stages with inter-stage cooling.

TABLE 19. TYPICAL SYNTHESIS GAS COMPOSITION BEFORE AND AFTER METHANATION (53)

Constituent	Composition, vol, % Before	After
H_2	74.5	74.1
N_2	24.3	24.7
CO	0.4	<5 ppm
CO_2	0.05	<5 ppm
CH_4	0.45	0.9
Ar	0.3	0.3
O_2	-	-

NH_3 Synthesis

The compressed synthesis gas is now suitable for use in the production of NH_3 and enters the NH_3 generating section of the plant. Since, even with high operating pressures and a favorable mol ratio of $N_2:H_2$, the equilibrium concentration of NH_3 is relatively low (13-17%), the system must be designed to allow the gases to make several passes through the catalytic reactor. As the gases leave the NH_3 synthesis reactor they are cooled in a heat exchanger and, in most cases, are passed through a refrigeration unit to condense most of the NH_3 produced. The resulting gas-liquid mixture is separated and the liquid NH_3 at -33°C (-28°F) is pumped to the storage tanks while most of the remaining synthesis gas is recycled to the NH_3 reactor. A small amount of this recycle gas is purged to prevent the buildup of inert gases in the loop and makeup synthesis gas is then added to the recycle mixture and passed through the catalytic reactor. Since the amount of NH3 formed and hence removed during each pass is so low, the actual gas stream circulating around the loop is very large in comparison with inlet and exit gas streams.

The catalytic reactor used to synthesize NH_3 typically operates at about 500-510°C (932-950°F) and has an intricate internal design. This type of reactor is required since the NH_3 reaction is highly exothermic but the catalyst activity and kinetics limit the acceptable temperatures in the reactor to a narrow range. That is, if the reactor temperatures increase, the catalyst activity significantly decreases but if the temperatures decrease, the rate of formation of NH3 significantly decreases. Conversion per pass decreases in either case.

To overcome these difficult process control problems, the so-called quench converter design, as shown in Figure 12, was developed. The catalyst particles typically contain 93% iron oxide (Fe_2O_3), 3.3% Al_2O_3, 0.67% magnesia (MgO), 0.55% silicone dioxide (SiO_2), 3.0% CaO, and 0.65% K_2O_3. This catalyst is separated into about three different beds with relatively "cold" synthesis gas added between each bed. This arrangement results in the partially converted

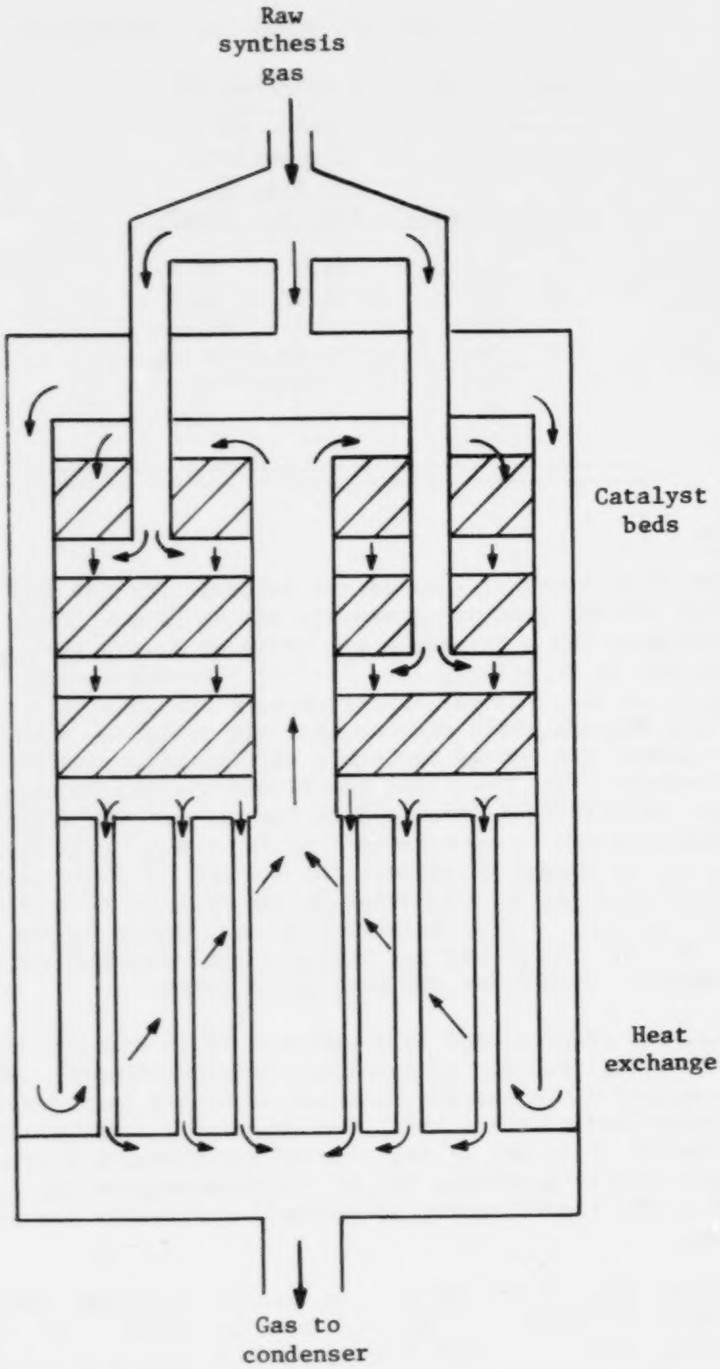

Figure 12. Diagram of quench-type converter (53).

gas being quenched as it leaves each catalyst bed. The remaining cool synthesis gas passes around the outside of the reactor and enters an internal heat exchanger to extract reaction heat from the converter product gas. Thus the incoming synthesis gas is preheated to the reaction temperature. The preheated gas then enters the catalyst beds and is partially converted to NH_3 before exiting the reactor through the internal heat exchanger.

Raw Material, Utility, Labor, and Energy Requirements

The raw material, utility, labor, and energy requirements to produce 1 ton of NH_3 in a 1000 ton/day NH_3 plant based on using natural gas as the H_2 feedstock are shown in Table 20. The major item under both raw materials and utilities is natural gas. In terms of the overall energy required to produce NH_3, natural gas contributes essentially all of the input energy.

TABLE 20. UNIT RAW MATERIAL, UTILITY, LABOR, AND ENERGY REQUIREMENTS FOR THE PRODUCTION OF NH_3 FROM NATURAL GAS[a] (47)

Item	Quantity required, unit/ton NH_3	Energy equivalent, MBtu/ton NH_3
Raw material		
Natural gas, kft^3	19.0	19.0
Utility		
Natural gas, kft^3	12.6	12.6
Electricity, kWh	30.0	0.1
Cooling water, kgal	50.0	-
Boiler feedwater, kgal	0.5	-
Labor, man-hr	0.18	-
Total		31.7

a. Basis
Plant capacity, 1000 ton/day.
On-stream time, 330 days/yr.

Process Economics

The total capital investment and the various raw material, utility, and labor requirements for a natural gas-based plant producing 1000 tons/day of NH_3 have been previously estimated by others (46) and were used as the basis for this study. However, the unit costs for each of the utilities except natural gas was modified to reflect updated projections for 1978 costs. Projecting the major raw material cost, i.e., the price of natural gas, is relatively uncertain at the present time and for this reason several alternate natural gas prices were chosen. Since the current price of natural gas ranges up to $2.50/MBtu and is dependent on many factors, $1.00/MBtu was chosen as the low price alternative and $3.00/MBtu was chosen as the middle range gas price. The higher natural gas price, $5/MBtu, was included primarily to serve as an upper limit on the raw material cost. The only other direct costs, maintenance, and analyses requirements were taken from data given by others (46). The major indirect cost, capital charges, was calculated at 22.1% of the total capital investment based on joint EPA-TVA premises. (A breakdown of this total capital charge is shown in Table 21.) The other indirect costs, the various overhead charges, were taken from a standard text on chemical engineering economics (48). The factors used for each of these indirect costs are shown in Table 22.

TABLE 21. ANNUAL CAPITAL CHARGES

FOR CHEMICAL INDUSTRY FINANCING

Item	Percent
Depreciation (straight line - 11 yr)	9.1
Insurance	0.5
Property taxes	1.5
Total rate applied to original investment	11.1
Cost of capital (capital structure assumed to be 40% debt and 60% equity)	
Bonds at 10% interest	4.0
Equity at 15% return to stockholder	9.0
Income taxes (Federal and State)[a]	9.0
Total rate applied to depreciation base	22.0[b]

a. Since income taxes are approximately 50% of gross return, the amount of taxes is the same as the return on equity.

b. Applied on an average basis, the total annual percentage of original fixed investment for new plants would be:

$$11.1\% + 1/2\ (22.0) = 22.1\%$$

TABLE 22. ASSUMPTIONS USED TO CALCULATE THE NH_3 PRODUCTION COSTS (48)

Item	Assumption
Direct Cost	
Maintenance	5% of capital investment
Indirect Cost	
Overhead	
Plant	50% of conversion costs less utilities
Administrative	15% of operating labor and maintenance
Selling costs	2% of direct and indirect costs
Research and development	2% of direct and indirect costs

The total capital investment for a 1000 ton/day NH_3 plant based on natural gas feedstock was estimated at $71.1 M in 1978 and the resulting unit NH_3 production cost ranges from about $111/ton for natural gas priced at $1/MBtu to $237/ton for natural gas at $5/MBtu. The detailed calculations for the $1/MBtu, $3/MBtu, and $5/MBtu cases are shown in Tables 23, 24, and 25 respectively.

The major cost associated with producing NH_3 from natural gas (for the $1/MBtu case) is the capital charges, representing approximately 43.0% of the total annual cost. The second largest annual expense is the cost of natural gas, about 28.5% of the total when the price of natural gas is $1/MBtu. This expense for natural gas can be further split into feedstock natural gas, making up 17.1% of the total annual operating cost, and natural gas used as fuel, which makes up the remaining 11.4% of the total annual operating cost. The other annual operating costs were each less than 10% of the total annual operating cost.

OTHER CURRENT NH_3 GENERATING TECHNIQUES

Since the major raw material for NH_3 production, other than N_2 which is readily and freely available from air, is H_2, any hydrocarbon or hydrocarbon mixture is a potential source of H_2. Although any hydrocarbon would be technically feasible, economics play a major role in specifying which, if any, H_2 source will replace natural gas.

Any other hydrocarbon source must generate sufficient economic savings, such as low purchase cost or uninterruptable supply, to overcome the two major advantages of natural gas, i.e., its high purity as received and its high H_2 to inert composition. Since natural gas from a pipeline already has most of the sulfur and other potential catalyst poisons removed, a large-scale gas purification system is not required. Any poisons which are present can

TABLE 23. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING NATURAL GAS AS THE H_2 FEEDSTOCK[a]

(LOW FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Natural gas	6,270,000 MBtu	1.00/MBtu	6,270,000	17.1
Catalyst and chemicals			225,000	0.6
Total raw materials costs			6,495,000	17.7
Conversion				
Operating labor and supervision	55,700 man-hr	10.00/man-hr	557,000	1.5
Utilities				
Natural gas	4,170,000 MBtu	1.00/MBtu	4,170,000	11.4
Electricity	9,900,000 kWh	0.027/kWh	267,000	0.7
Cooling water	16,500,000 kgal	0.11/kgal	1,815,000	5.0
Boiler feedwater	180,000 kgal	0.92/kgal	166,000	0.5
Maintenance, 5% of plant investment			3,325,000	9.1
Analyses	2,080 man-hr	15.00/man-hr	31,000	0.1
Total conversion costs			10,331,000	28.3
Total direct costs			16,826,000	46.0
Indirect Costs				
Capital charges, 22.1% of capital investment			15,713,000	43.0
Overhead				
Plant, 50% of conversion costs less utilities			1,957,000	5.4
Administrative, 15% of operating labor and maintenance			587,000	1.6
Marketing, 2% of direct and indirect costs			731,000	2.0
Research and development, 2% of direct and indirect costs			731,000	2.0
Total indirect costs			19,719,000	54.0
Total annual operating costs			36,545,000	100.0
	$/ton of NH_3			
Equivalent unit operating costs	110.74			

a. Basis (47)
Total capital investment, $71,100,000.
Total plant investment, $66,500,000.
On-stream time, 330 days/yr.

TABLE 24. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING NATURAL GAS AS THE H_2 FEEDSTOCK[a]

(MEDIUM FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Natural gas	6,270,000 MBtu	3.00/MBtu	18,810,000	32.7
Catalyst and chemicals			225,000	0.4
Total raw materials costs			19,035,000	33.1
Conversion				
Operating labor and supervision	55,700 man-hr	10.00/man-hr	557,000	1.0
Utilities				
Natural gas	4,170,000 MBtu	3.00/MBtu	12,510,000	21.7
Electricity	9,900,000 kWh	0.027/kWh	267,000	0.5
Cooling water	16,500,000 kgal	0.11/kgal	1,815,000	3.2
Boiler feedwater	180,000 kgal	0.92/kgal	166,000	0.3
Maintenance, 5% of plant investment			3,325,000	5.7
Analyses	2,080 man-hr	15.00/man-hr	31,000	0.1
Total conversion costs			18,671,000	32.5
Total direct costs			37,706,000	65.6
Indirect Costs				
Capital charges, 22.1% of capital investment			15,713,000	27.4
Overhead				
Plant, 50% of conversion costs less utilities			1,957,000	3.4
Administrative, 15% of operating labor and maintenance			587,000	1.0
Marketing			731,000	1.3
Research and development			731,000	1.3
Total indirect costs			19,719,000	34.4
Total annual operating costs			57,425,000	100.0
	$/ton of NH_3			
Equivalent unit operating costs	174.02			

[a] Basis (47)
Total capital investment, $71,100,000.
Total plant investment, $66,500,000.
On-stream time, 330 days/yr.

TABLE 25. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING NATURAL GAS AS THE H_2 FEEDSTOCK[a]

(HIGH FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Natural gas	6,270,000 MBtu	5.00/MBtu	31,350,000	40.0
Catalyst and chemicals			225,000	0.3
Total raw materials costs			31,575,000	40.3
Conversion				
Operating labor and supervision	55,700 man-hr	10.00/man-hr	557,000	0.7
Utilities				
Natural gas	4,170,000 MBtu	5.00/MBtu	20,850,000	26.6
Electricity	9,900,000 kWh	0.027/kWh	267,000	0.3
Cooling water	16,500,000 kgal	0.11/kgal	1,815,000	2.3
Boiler feedwater	180,000 kgal	0.92/kgal	166,000	0.2
Maintenance, 5% of plant investment			3,325,000	4.2
Analyses	2,080 man-hr	15.00/man-hr	31,000	[b]
Total conversion costs			27,011,000	34.5
Total direct costs			58,586,000	74.8
Indirect Costs				
Capital charges, 22.1% of capital investment			15,713,000	20.1
Overhead				
Plant, 50% of conversion costs less utilities			1,957,000	2.5
Administrative, 15% of operating labor and maintenance			587,000	0.8
Marketing			731,000	0.9
Research and development			731,000	0.9
Total indirect costs			19,719,000	25.2
Total annual operating costs			78,305,000	100.0
Equivalent unit operating costs	$/ton of NH_3 237.29			

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a. Basis (47)
Total capital investment, $71,100,000.
Total plant investment, $66,500,000.
On-stream time, 330 days/yr.

b. Less than 0.1%.

be removed with the installation of guard boxes before the catalytic reactors. Natural gas also primarily contains methane which contains a higher hydrogen to carbon ratio than any other hydrocarbon.

For these reasons, it is not surprising that about 95% of the current NH_3 production in the United States is generated from natural gas. The remaining 5% of the NH_3 production is, in most instances, special cases where site-specific circumstances provide favorable economics. These miscellaneous sources include refinery gas, the off-gas from electrolytic cells, and coke-oven gas.

Refinery Gas

Refinery gas streams would seem to be a logical raw material for NH_3 since refinery operations involve the treatment of various hydrocarbon streams. Of the many potential off-gas streams in a refinery, the two streams with the most potential as a feedstock are the gases from the catalytic cracking units and the catalytic reformer. Typical compositions of these two refinery streams are shown in Table 26. Of these two streams, the off-gas from the catalytic reformer shows the most potential since it has a higher H_2 concentration and no O_2-containing compounds. For use in an NH_3 plant the only pretreatment step required would be stepwise refrigeration to remove the various hydrocarbon gases, leaving a high purity H_2 stream. The off-gas from the catalytic cracking units, on the other hand, contains all of the impurities present in the reformer off-gas and in addition contains O_2, CO, and CO_2 which must be removed.

TABLE 26. TYPICAL COMPOSITIONS OF REFINERY OFF-GASES (53)

Component	Gas from a catalytic cracker, vol %	Gas from a catalytic reformer, vol %
H_2	34.2	62.0
CO_2	1.3	-
CO	1.2	-
O_2	0.1	-
N_2	5.5	-
C_1	19.2	17.2
C_2	23.5	9.7
C_3	12.7	4.5
C_4	2.6	0.4
C_5	0.7	0.2
Sulfur	1,200[a]	-

[a] Gr/100 ft^3.

The primary advantages of using refinery gas streams are the relatively low cost and availability of these H_2-rich streams. In fact several oil companies entered the NH_3 market in the late 1960's by adding NH_3 synthesis loops at their refineries. However, recently the requirements for H_2 within the refinery for both hydrocracking and hydrotreating, particularly hydrodesulfurization, have increased such that some refineries have begun to build gas reformers to generate additional H_2. Thus the future potential for generating NH_3 from refinery gas streams appears to be very limited.

Electrolytic Cells

Another source of H_2 for NH_3 production is the off-gas from electrolytic cells. Cl_2 gas is typically generated by passing an electric current through a nearly saturated solution of sodium chloride (NaCl) at 60-70°C (140-158°F). The Cl_2 is removed as a gas at the anode and byproduct H_2 gas is removed at the cathode. This H_2 can then be simply mixed with N_2, compressed, and sent to the NH_3 synthesis loop. This type of process has the economic advantages of (1) converting a waste gas stream to a marketable byproduct and (2) at the same time eliminating some of the processing steps usually required in the production of NH_3 from natural gas (reforming and synthesis gas purification). The major problems associated with using H_2 from electrolytic Cl_2 cells to produce NH_3 are the small-scale demand for Cl_2 compared with the yearly demand for NH_3 and also, in a related matter, the small number of large-scale Cl_2 generating plants with the capacity to supply H_2 to economically justify the construction of a large NH_3 plant. For example, a 500 ton/day Cl_2 plant produces enough H_2 for about 60 tons/day NH_3. Thus the potential for producing NH_3 from electrolytic cell off-gas in the future is severely limited.

At the present time, only six NH_3 plants in the United States are currently using H_2 from electrolytic Cl_2 cells and these plants have a total capacity of about 315,000 tons/yr (with a range of 8,000 to 115,000 tons/yr). When compared with the standard natural gas-based NH_3 plants with an annual capacity of 400,000 tons/yr for a single plant, they are relatively small and represent only a minor portion of the total U.S. NH_3 capacity.

Coke-Oven Gas

The only other current alternative source of NH_3 in the United States is coke-oven gas. Before the advent of low cost, readily available natural gas, coke-oven gas was a major source of NH_3. However, only two major plants are now operating and their total output (470,000 tons/yr) represents only about 1.5% of the current U.S. NH_3 production. These plants are located at U.S. Steel's Geneva, Utah, and Clairton, Pennsylvania, coke ovens. The Clairton plant is the larger one producing approximately 1200 tons/day of NH_3.

A quick analysis of the method of producing coke demonstrates why coke-oven gas was a major source of NH_3 in the past. The initial charge of crushed coal is fed to a coproduct coke oven, a refractory-lined, gas-tight bin with typical dimensions of 40 x 1.5 x 13 ft. These individual ovens are arranged in batteries of 10-100 ovens. As heat and air are added to the oven, the volatile matter in the coal, i.e., the lower molecular weight hydrocarbons, the pyrolyzed hydrocarbons, and some of the impurities in the coal are driven off as gases.

As shown in the block flow diagram in Figure 13, the gas leaving the ovens at about 1250-1300°C (2282-2372°F) is initially contacted with a weak aqueous NH_3 spray which removes most of NH_3 and the tars from the gas. In the process the gas temperature is reduced to about 100°C (212°F). The liquid mixture is separated by density into an NH_3 solution and byproduct tars. The remaining tar is removed from the gas by an electrostatic precipitator (ESP) before the gas is scrubbed with sulfuric acid (H_2SO_4) to remove any remaining NH_3. After a final scrubbing stage of oil is used to remove the light aromatics such as benzene, the "partially cleaned" coke-oven gas leaves this cleanup section of the coke-oven battery with the typical composition shown in Table 27. The major constituents of the gas, H_2 and methane, are usually the two prime H_2 feedstocks for NH_3 production. However, in most instances during the further purification of the gas for the production of NH_3, all the remaining constituents except H_2 are removed by cryogenic separation. These higher boiling point impurities, primarily hydrocarbon gases, are used as boiler fuel and the remaining N_2-H_2 stream is then compressed and sent to the NH_3 synthesis loop.

TABLE 27. TYPICAL COMPOSITION OF COKE-OVEN GAS (2, 53)

Constituent	Vol, %
H_2	4.7
N_2	6.6
CO	7.5
O_2	0.6
CH_4	25.1
C_2H_4	1.8
$C_2H_6 + C_3H_8$	0.4
$C_3H_6 + C_4H_8$	0.4
$C_6H_6 + C_2H_2$	0.2
CO_2	2.1
H_2S	1 gr/Nm^3
NO	Trace
HCN	5.2 gr/Nm^3

The major disadvantage in using coke-oven gas for the production of NH_3 is the need for an extensive purification system. This is, however, partially balanced by the fact that some of the impurities must be removed to meet air pollution regulations and most of the remaining impurities which are removed can be either sold as a byproduct [e.g., tars, aromatics, ammonium sulfate [$(NH_4)_2SO_4$] or used within the plant as boiler fuel. Another disadvantage is that a large number of individual ovens must be located in a single battery to generate sufficient gas to justify building both a purification system and an NH_3 plant, particularly since the coke-oven gas can be used as a boiler fuel without expending the additional capital for processing the gas to NH_3. This

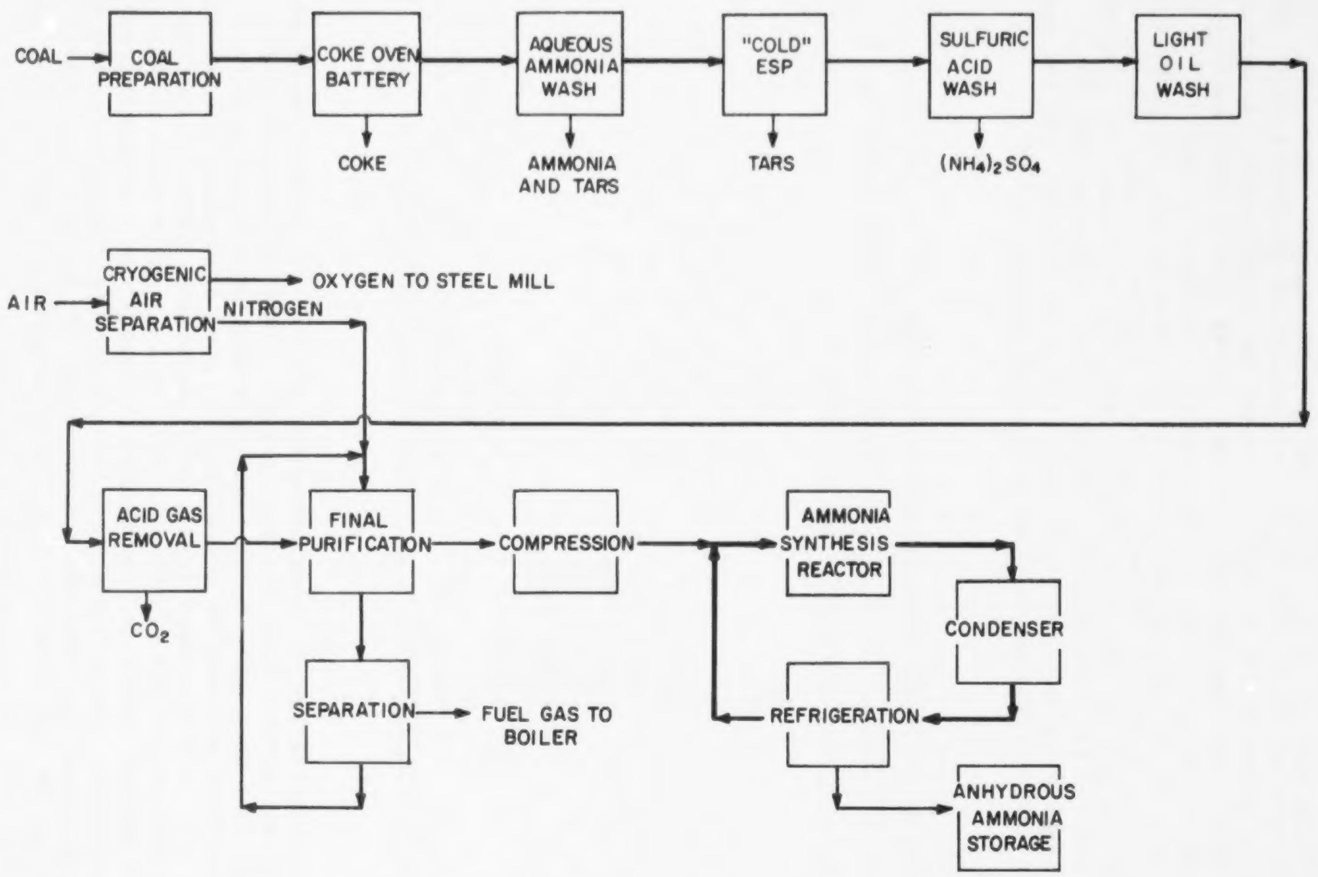

Figure 13. Block flow diagram of a typical NH_3 plant based on coke-oven gas as the H_2 feedstock.

last disadvantage implies that this type of NH_3 plant will be located only where extensive steel-making plants are being built and hence has limited potential for large-scale production of NH_3 in the future.

The primary advantage of using coke-oven gas as a raw material for NH_3 is that coke-oven gas is a byproduct of steel making and the only cost associated with its purchase is the expense of replacing its heating value as a boiler fuel. Another advantage is the extreme high purity of the H_2 stream after treatment, the level of inerts in the NH_3 synthesis loop is low enough that the usual purge stream is not required.

Raw Material, Utility, Labor, and Energy Requirements

The raw material, utility, labor, and energy requirements for the production of 1 ton of NH_3 from coke-oven gas are given in Table 28.

TABLE 28. UNIT RAW MATERIAL, UTILITY, LABOR, AND

ENERGY REQUIREMENTS FOR THE PRODUCTION OF NH_3 FROM COKE-OVEN GAS (53)

Item	Quantity required, unit/ton NH_3	Energy equivalent, MBtu/ton NH_3
Raw material		
Coke-oven gas, kft^3	147.6	60.0
N_2, kft^3	29.8	-
Utility		
Fuel, kft^3	(61.0)[a]	(38.7)[a]
Electricity, kWh	730	6.5
Cooling water, kgal	24	-
Steam, MBtu	0.95	1.0
Labor, man-hr	-	-
Total		28.8

a. Surplus fuel gas.

POTENTIAL FUTURE NH_3 GENERATING TECHNIQUES

STEAM REFORMING OF NAPHTHA (27, 53)

Although not used in the United States at the present time, another potentially significant raw material, which is widely used in areas of the world (Great Britain, India, and Japan) where natural gas is not readily available, is naphtha. Naphtha is a light petroleum distillate (maximum boiling point, 215°C) which is commercially available and can be reformed in a similar manner as natural gas. The main problems are to prevent carbon deposition (coking) on the catalyst and to control excessive steam consumption during reforming. The primary disadvantage when using naphtha as a raw material feedstock is its desirability as a feedstock for other petrochemical operations and gasoline blending. This latter use, at least for the U.S. market in the near future, eliminates its potential for NH_3 production since its value in gasoline is significantly greater than its value as a feedstock for NH_3 production. In addition to its higher cost, naphtha also has a much higher sulfur content than natural gas and hence requires a purification system to remove these sulfur compounds.

Process Description

The first step in processing naphtha to NH_3, as shown in Figure 14, is a two-stage hydrodesulfurization system in which most of the sulfur in the naphtha is converted to H_2S which is then adsorbed by hot Fe_2O_3. This purification step usually involves a two-stage catalytic hydrogenation in which the naphtha is vaporized and mixed with an H_2-rich gas. As this mixture passes over the catalyst, the sulfur present in the naphtha is converted to H_2S. During the regeneration of the iron sulfide (FeS) the sulfur is reconverted to sulfur dioxide (SO_2) and sent as a concentrated stream to a Claus unit.

The partially cleaned naphtha is mixed with steam and sent to a primary reformer similar to that used in a conventional natural gas-based NH_3 plant. The operating temperature in the primary reformer is about 750-850°C (1382-1562°F) at a pressure of 20-30 atmosphere and a steam:carbon mol ratio ranging from 2.0 to 3.0:1. The catalyst used in the primary reformer is Ni-based; however, it also contains small amounts of potassium oxide (K_2O). The catalyst support material is typically kalsilite ($K_2O \cdot Al_2O_3 \cdot 2SiO_2$) and the catalyst particles are usually ring shaped.

The remaining sections of the naphtha-based NH_3 plants are identical with those previously shown for an NH_3 plant using natural gas as the H_2 feedstock. The partially reformed gas is mixed with air and passed through a secondary reformer to complete the conversion of naphtha to CO and H_2. The

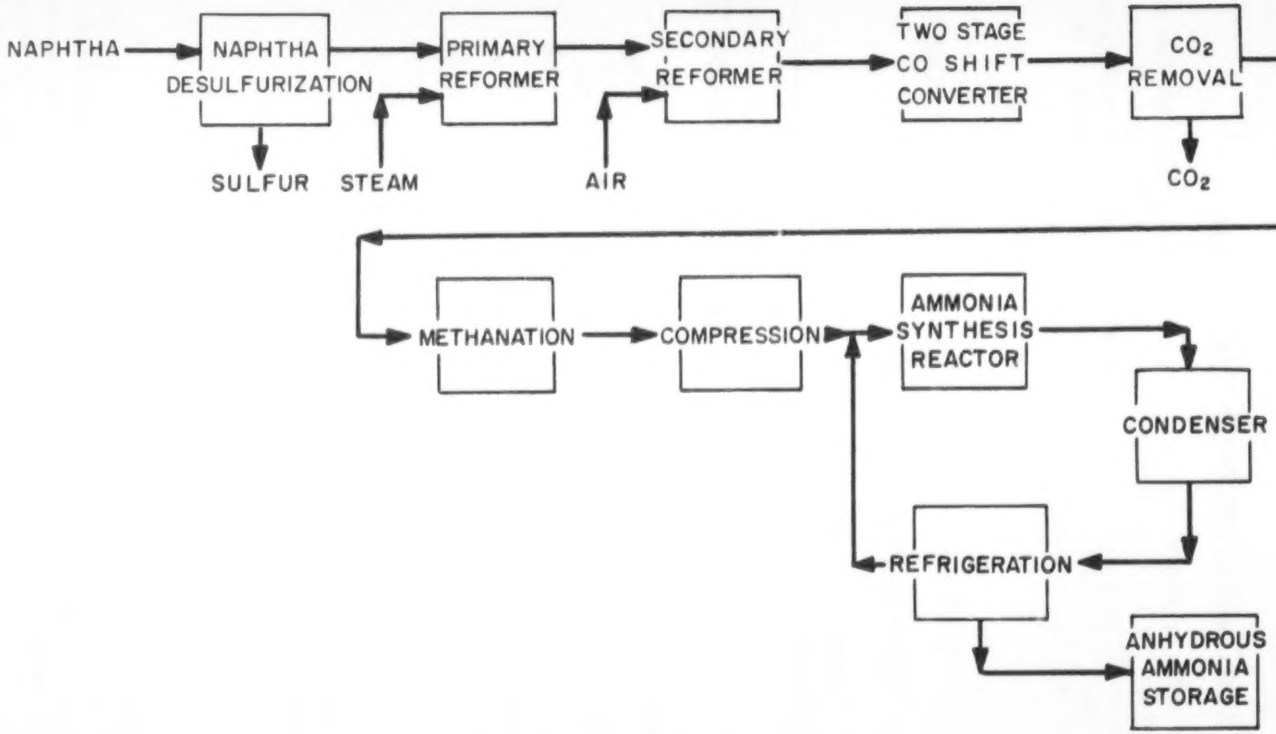

Figure 14. Block flow diagram of a typical NH_3 plant based on naphtha as the H_2 feedstock.

CO is converted to CO_2 in a two-stage water gas shift reaction system to maximize the H_2 production and the resulting CO_2 (as well as the CO_2 formed during the reforming operation) is removed by a "promoted" K_2CO_3 system. As before, the remaining small quantities of CO are destroyed in a methanation reactor and the resulting clean synthesis gas is compressed and sent to the NH_3 synthesis loop.

Raw Material, Utility, Labor, and Energy Requirements

The raw material, utility, labor, and energy requirements for producing 1 ton of NH_3 by steam reforming naphtha in a 1000 ton/day plant are shown in Table 29. The major item under both raw material and utility requirements is naphtha. In terms of the overall energy required to generate NH_3, naphtha contributes essentially all (99.5%) of the input energy.

TABLE 29. UNIT RAW MATERIAL, UTILITY, LABOR, AND ENERGY REQUIREMENTS FOR THE PRODUCTION OF NH_3 FROM NAPHTHA BY STEAM REFORMING[a] (47)

Item	Quantity required, unit/ton NH_3	Energy equivalent, MBtu/ton NH_3
Raw material		
Naphtha, MBtu	19.4	19.4
Utility		
Naphtha, MBtu	14.1	14.1
Electricity, kWh	45.5	0.2
Cooling water, kgal	65.5	-
Boiler feedwater, kgal	0.5	-
Labor, man-hr	0.20	-
Total		33.7

a. Basis
Plant capacity, 1000 tons/day.
On-stream time, 330 days/yr.

Process Economics

The total capital investment and the various raw material, utility, and labor requirements for a naphtha-based NH_3 plant producing 1000 tons/day of NH_3 have been previously estimated by others (47). In a similar manner as was used in pricing natural gas, the utility and labor costs were based on updated 1978 costs. Since naphtha is an oil-based petrochemical and hence may be subjected to higher than average escalation rates in the future, three alternate costs were assumed for the naphtha. The lowest price assumed for naphtha in 1978 was $0.40/gal, its current price. The middle price and the high price alternatives were chosen as $0.60 and $0.80/gal, respectively, reflecting higher feedstock prices. The other direct and the various indirect costs were calculated in the same manner as was previously discussed for the natural gas-based NH_3 plant.

The total capital investment for a 1000 ton/day NH_3 plant based on steam reforming naphtha was estimated as $82.4 M and the resulting unit production cost for NH_3 ranged about $193/ton for naphtha at $0.40/gal to about $294/ton for naphtha at $0.80/gal. The detailed calculation for the $0.40/gal naphtha case is shown in Table 30 and the detailed calculations for the other two cases are given in Table 31 and 32.

The major cost associated with producing NH_3 from naphtha (for the $0.40/gal case) is the naphtha cost, which accounts for about 51.9% of the annual production cost. The feedstock naphtha accounts for 30.0% of the annual operating cost with the naphtha used as fuel contributing the remaining 21.9%. The capital charges, which are the most important annual cost in the natural gas-based plant, represent about 28.6% of the total annual operating cost with the other various direct and indirect expenses each contributing less than 6% of the total annual cost.

PARTIAL OXIDATION OF HYDROCARBONS (53)

As the heavier hydrocarbons above methane are used as feedstocks for NH_3 production, the conversion of these hydrocarbons to H_2 by steam reforming becomes more difficult and more unattractive economically. Not only are the sulfur content and the content of other impurities higher, thus requiring larger purification systems (and thus substantially larger amounts of H_2) but also the hydrogen to carbon ratio decreases in the heavier petroleum fractions. A declining hydrogen to carbon ratio requires an increasing steam to hydrocarbon ratio to achieve steam reforming without the deposition of carbon on the reforming catalyst. Usually naphtha is the heaviest petroleum fraction reformed with steam because of the large quantities of steam required and the operating problems associated with reforming higher hydrocarbon streams.

Further complicating this hydrocarbon feedstock problem, the lower molecular weight petroleum fractions are more desirable and hence more expensive than the heavy hydrocarbons, such as heavy fuel oil or residual oils. Technically any petroleum fraction can be used in this system; fuel oil and naphtha are the two fractions most commonly mentioned. Since naphtha has already been discussed under steam reforming, only heavy fuel oil will be considered here.

Although not economically attractive where natural gas is readily available, considerable work has recently been done on a method of converting these heavier petroleum fractions into H_2 and CO by partial oxidation. Because of environmental regulations, they can no longer be burned without hydrotreating the fuel or the installation of FGD systems. A typical overall processing scheme for partial oxidation based on a heavy oil feedstock is shown in Figure 15.

TABLE 30. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING NAPHTHA (STEAM REFORMING) AS THE H_2 FEEDSTOCK[a]

(LOW FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Naphtha	6,395,000 MBtu	3.00/MBtu	19,185,000	30.0
Catalyst and chemicals			300,000	0.5
Total raw materials costs			19,485,000	30.5
Conversion				
Operating labor and supervision	62,300 man-hr	10.00/man-hr	623,000	1.0
Utilities				
Naphtha	4,640,000 MBtu	3.00/MBtu	13,920,000	21.9
Electricity	15,000,000 kWh	0.027/kWh	405,000	0.6
Cooling water	21,600,000 kgal	0.11/kgal	2,376,000	3.7
Boiler feedwater	150,000 kgal	0.92/kgal	138,000	0.2
Maintenance, 5% of plant investment			3,755,000	5.9
Analyses	2,080 man-hr	15.00/man-hr	31,000	-[b]
Total conversion costs			21,248,000	33.3
Total direct costs			40,733,000	63.8
Indirect Costs				
Capital charges, 22.1% of capital investment			18,210,000	28.6
Overhead				
Plant, 50% of conversion costs less utilities			2,205,000	3.5
Administrative, 15% of operating labor and maintenance			661,000	1.0
Marketing			731,000	1.1
Research and development, 2% of direct and indirect costs			1,276,000	2.0
Total indirect costs			23,083,000	36.2
Total annual operating costs			63,816,000	100.0
	$/ton of NH_3			
Equivalent unit operating costs	193.38			

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a. Basis (47)
Total capital investment, $82,400,000.
Total plant investment, $75,100,000.
On-stream time, 330 days/yr.

b. Less than 0.1%.

TABLE 31. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING NAPHTHA (STEAM REFORMING) AS THE H_2 FEEDSTOCK[a]

(MEDIUM FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Naphtha	6,395,000 MBtu	4.50/MBtu	28,778,000	35.8
Catalyst and chemicals			300,000	0.4
Total raw materials costs			29,078,000	36.2
Conversion				
Operating labor and supervision	62,300 man-hr	10.00/man-hr	623,000	0.8
Utilities				
Naphtha	4,640,000 MBtu	4.50/MBtu	20,880,000	26.0
Electricity	15,000,000 kWh	0.027/kWh	405,000	0.5
Cooling water	21,600,000 kgal	0.11/kgal	2,376,000	2.9
Boiler feedwater	150,000 kgal	0.92/kgal	138,000	0.2
Maintenance, 5% of plant investment			3,755,000	4.7
Analyses	2,080 man-hr	15.00/man-hr	31,000	–[b]
Total conversion costs			28,208,000	35.1
Total direct costs			57,286,000	71.3
Indirect Costs				
Capital charges, 22.1% of capital investment			18,210,000	22.7
Overhead				
Plant, 50% of conversion costs less utilities			2,205,000	2.7
Administrative, 15% of operating labor and maintenance			661,000	0.8
Marketing			731,000	0.9
Research and development			1,276,000	1.6
Total indirect costs			23,083,000	28.7
Total annual operating costs			80,369,000	100.0
	$/ton of NH_3			
Equivalent unit operating costs	243.54			

[a] Basis (47)
Total capital investment, $82,400,000.
Total plant investment, $75,100,000.
On-stream time, 330 days/yr.

[b] Less than 0.1%.

TABLE 32. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING NAPHTHA (STEAM REFORMING) AS THE H_2 FEEDSTOCK[a]

(HIGH FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Naphtha	6,395,000 MBtu	6.00/MBtu	38,370,000	39.6
Catalyst and chemicals			300,000	0.3
Total raw materials costs			38,670,000	39.9
Conversion				
Operating labor and supervision	62,300 man-hr	10.00/man-hr	623,000	0.7
Utilities				
Naphtha	4,640,000 MBtu	6.00/MBtu	27,840,000	28.7
Electricity	15,000,000 kWh	0.027/kWh	405,000	0.4
Cooling water	21,600,000 kgal	0.11/kgal	2,376,000	2.5
Boiler feedwater	150,000 kgal	0.92/kgal	138,000	0.1
Maintenance, 5% of plant investment			3,755,000	3.9
Analyses	2,080 man-hr	15.00/man-hr	31,000	–[b]
Total conversion costs			35,168,000	36.3
Total direct costs			73,838,000	76.2
Indirect Costs				
Capital charges, 22.1% of capital investment			18,210,000	18.8
Overhead				
Plant, 50% of conversion costs less utilities			2,205,000	2.3
Administrative, 15% of operating labor and maintenance			661,000	0.7
Marketing			731,000	0.7
Research and development			1,276,000	1.3
Total indirect costs			23,083,000	23.8
Total annual operating costs			96,921,000	100.0
		$/ton of NH_3		
Equivalent unit operating costs		293.70		

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a. Basis (47)
Total capital investment, $82,400,000.
Total plant investment, $75,100,000.
On-stream time, 330 days/yr.

b. Less than 0.1%.

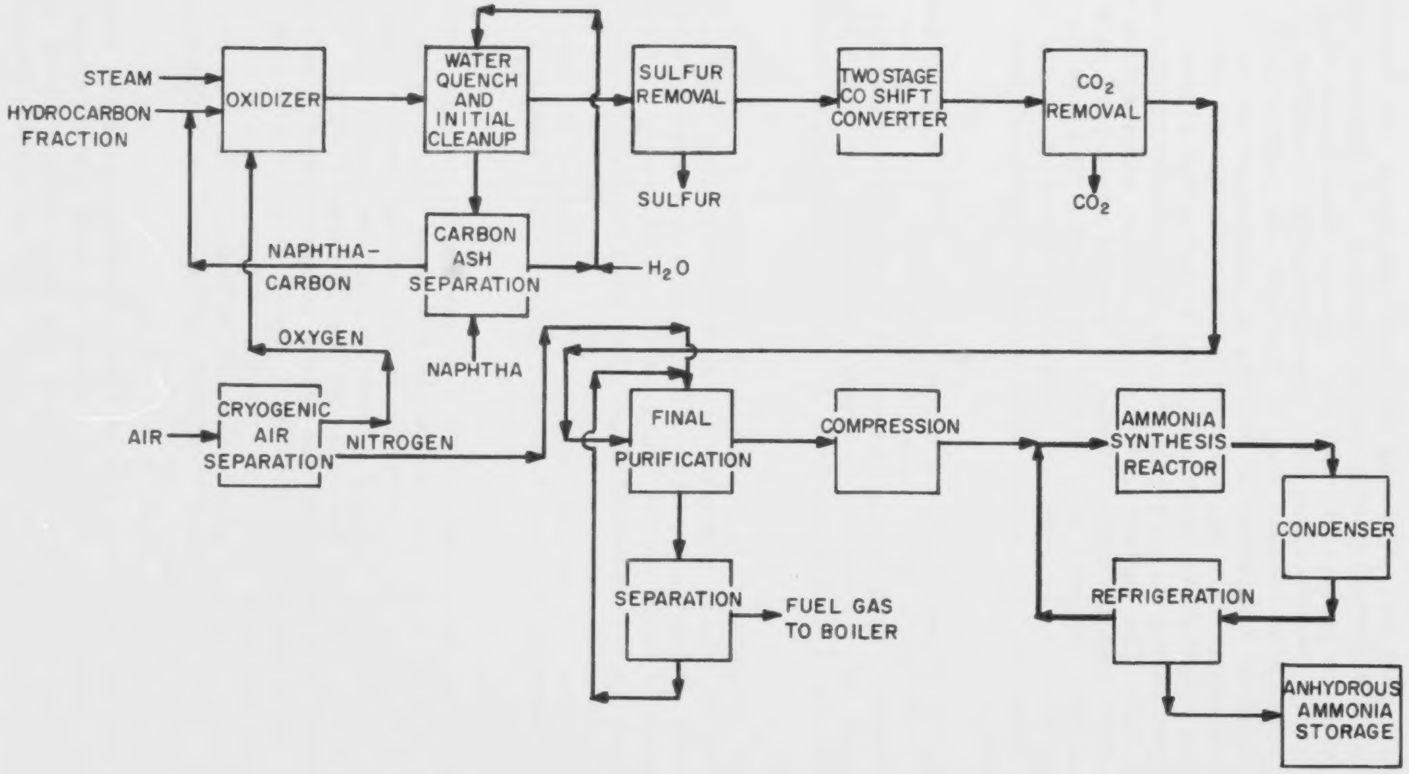

Figure 15. Block flow diagram of a typical NH3 plant based on heavier hydrocarbons as the H_2 feedstock.

Process Description

Partial oxidation involves injecting the heavy petroleum fraction and O_2, although less O_2 than the stoichiometric amount required for complete combustion, into a reaction chamber to generate H_2 and CO. In some cases, described later, steam is also injected. The resulting hot, gas-containing impurities are quenched with H_2O, which also removes most of the soot and particulates. The gas is further purified in a scrubber by passing it countercurrent to an H_2O wash stream. The wastewater from the scrubber becomes the quench water and is eventually sent to a carbon recovery system as a carbon-H_2O slurry. The carbon is recovered by mixing naphtha with slurry which preferentially separates the carbon from the H_2O. The naphtha-carbon layer is then removed from the top of a decanter. The naphtha-carbon stream is usually mixed with part of the hydrocarbon feedstock and sent to a distillation unit. The naphtha is removed overhead and recycled to treat the carbon-H_2O slurry while the carbon-feedstock mixture can be used for either preheat fuel or recycled feedstock for partial oxidation.

When the feedstock is made up of saturated hydrocarbons, such as natural gas or other light hydrocarbon gases, steam injection into the reaction chamber is not required. However, when heavier petroleum feedstocks, containing olefins, such as heavy fuel are used, steam must be added to prevent excessive formation of carbon and hold carbon formation to 1-4% of the incoming carbon in the feedstock.

The product gas composition is dependent on many factors ranging from the pressure and temperature in the generator to the composition of the feedstock. Table 33 lists typical operating conditions and product gas compositions of three types of feedstocks based on the Shell Gasification Process. Table 34 gives the characteristics and compositions of the heavy fuel oil, light naphtha, and natural gas used in this gasification process. Since natural gas can be relatively easily reformed, it was included in this discussion for comparison purposes only; the remaining sections refer primarily to the partial oxidation of heavy fuel oil.

The first step after gas production is a sulfur removal stage. Although most of the sulfur is present in the raw gas as H_2S about 5% exists as carbonyl sulfide (COS). This further complicates the sulfur removal system since both H_2S and COS must be effectively removed. Various acid gas removal systems have been used including Alkazid, Sulfinol, or hot "promoted" K_2CO_3. Regardless of which of these methods is used, the gas passes countercurrently to the absorbing solution which removes H_2S and CO_2. The crude gas containing COS then typically passes through a catalytic reactor to hydrolyze the COS to H_2S and CO_2 and then passes through a second absorption tower. The sour off-gas from the absorption system containing about 15% H_2S is sent to a Claus unit to produce byproduct elemental sulfur.

The partially cleaned synthesis gas is then preheated, injected with steam, and passed through a conventional hot Fe_2O_3 catalyst where the water gas shift reaction occurs. The major difference between the shift reactors used in a conventional NH_3 plant (see Water Gas Shift section in Natural Gas section)

TABLE 33. TYPICAL OPERATING PARAMETERS FOR PRODUCTION OF NH_3 SYNTHESIS GAS FROM VARIOUS HYDROGEN FEEDSTOCKS USING THE SHELL GASIFICATION (PARTIAL OXIDATION) PROCESS (53)

	Feedstock[a]		
Parameter	Natural gas	Naphtha	Heavy fuel oil
Operating conditions			
Pressure, psig	450	450	450
Preheat temperatures, °F			
Feedstock	457	77	457
O_2	457	457	457
Steam	475	475	475
Temperature of crude gas, °F	285	95	95
Raw material consumption and utilities used per pound of feedstock			
O_2, sft^3	16.6	14.9	12.7
Steam, pound	0.05	0.35	0.40
Cooling water, gal	0.14	3.85	4.03
Boiler feedwater, gal	0.39	0.30	0.28
Electricity, kWh	7.6	11.4	11.4
Product gas per pound of feedstock			
Product gas, sft^3	63.2	54.3	49.9
($CO + H_2$) produced, sft^3	60.1	50.8	46.6
Composition, vol %			
H_2	60.9	51.6	46.1
CO	34.5	41.8	46.9
CO_2	2.8	4.8	4.3
CH_4	0.4	0.4	0.4
N_2	1.4	1.4	1.4
H_2S	-	0.007	0.9

a. Typical compositions are shown in Table 34.

TABLE 34. TYPICAL FEEDSTOCKS USED IN THE SHELL GASIFICATION PROCESS (PARTIAL OXIDATION) (53)

Natural Gas	
Characteristic	
Heating value, Btu/sft^3	937
Composition, vol %	
CH_4	97.31
C_2H_6	2.30
C_3H_8	0.20
C_4H_{10}	0.15
N_2	0.04
Light Naphtha	
Characteristics	
Specific gravity, 60/60	0.669
ASTM distillation, IB2, °F	99
ASTM distillation, FBP, °F	246
Heating value, Btu/lb	18,630
Composition, wt %	
Carbon	84.00
Hydrogen	15.97
Sulfur	0.03
Paraffins	86
Aromatics	5
Naphthenes	9
Heavy Fuel Oil	
Characteristics	
Specific gravity, 60/60	0.97
Viscosity, seconds Saybolt Universal at 100°F	3,500
Conradson carbon, wt%	10.2
ASTM pour point, °F	59
Heating value, Btu/lb	17,710
Composition, wt %	
Carbon	84.60
Hydrogen	11.30
Sulfur	3.50
Nitrogen	0.40
Ar	0.13
Ash	0.07

and those used in a partial oxidation system is that only a high-temperature catalyst is normally used and extra steam injectors are available if the inlet gas temperature is too low. The CO and H_2O react as follows:

$$CO_{(g)} + H_2O_{(g)} \rightarrow CO_{2(g)} + H_{2(g)} \qquad (9)$$

The CO_2 is then removed using conventional technology, such as MEA, TEA-MEA, or hot carbonate which were previously discussed.

Before introducing the synthesis gas to a liquid N_2 wash, the gas usually passes through a cold methanol wash (Rectisol) ro remove CO_2 and H_2O which would tend to form solids in the precooler at these cryogenic temperatures. Failure to remove these impurities would lead to frequent shutdowns of the cryogenic unit for defrosting.

The final purification step is usually a cold N_2 wash since cryogenic liquid N_2 is available as a byproduct from the air separation plant. The inlet synthesis gas is precooled by passing it countercurrently to the purified synthesis gas from the liquid N_2 scrubber. The cooled synthesis gas is then passed countercurrently to the liquid N_2. At these cryogenic temperatures the remaining impurities condense in the liquid N_2 and separate from the synthesis gas. In addition to the purifying the synthesis gas, the evaporation of liquid N_2 in the washing scrubber provides the N_2 to yield the H_2:N_2 mol ratio required for NH_3 production. The waste liquid N_2 from the washing scrubber is vaporized by passing it through an expansion valve and is used to liquefy the incoming N_2 gas from the air separation plant. The resulting waste gas as shown in Table 35 is used as boiler fuel.

TABLE 35. TYPICAL COMPOSITIONS OF GASES

TO AND FROM A CRYOGENIC PURIFICATION UNIT (53)

	Composition, vol %		
Constituent	Feed gas	Purified synthesis gas	Waste gas
H_2	94.0	75.0	16.0
N_2	0.1	25.0	44.0
CO	4.8	<5 ppm	33.0
CH_4	0.5	<1 ppm	3.5
Ar	0.1	<60 ppm	3.3
CO_2	0.001	-	-
H_2O	0.0003	-	-
O_2	0.0005	<10 ppm	-

The cleaned synthesis gas then passes through the precooler to cool the incoming synthesis gas before entering the synthesis loop. The impurities are reduced to such a low level, as shown in Table 35. that the purge stream from the synthesis loop is not needed; the few impurities that remain are dissolved in and removed with liquid NH_3 product.

Raw Material, Utility, Labor, and Energy Requirements

The raw material, utility, labor, and energy requirements for the production of 1 ton of NH_3 from heavy fuel oil in a 1000 ton/day NH_3 plant are given in Table 36. As in all the other NH_3 generating systems, the major item is the H_2 feedstock material since it is used both as a raw material and as a fuel in the system. The energy input in the fuel oil represents about 99.5% of the total energy consumed in the process.

TABLE 36. UNIT RAW MATERIAL, UTILITY, LABOR, AND ENERGY REQUIREMENTS FOR THE PRODUCTION OF NH_3 FROM HEAVY FUEL OIL[a] (47)

Item	Quantity required, unit/ton NH_3	Energy equivalent, MBtu/ton NH_3
Raw material		
Heavy fuel oil, MBtu	20.3	20.3
Utility		
Heavy fuel oil, MBtu	14.6	14.6
Electricity, kWh	136.0	0.5
Cooling water, kgal	80.0	-
Boiler feedwater, kgal	0.4	-
Labor, man-hr	0.23	-
Total		35.4

a. Basis
Plant capacity, 1000 tons/day.
On-stream time, 330 days/yr.

Process Economics

The total capital investment and the various raw material, utility, and labor requirements for a 1000 ton/day NH_3 plant based on the partial oxidation of heavy fuel oil have been previously estimated by others (47). The various factors used to calculate the indirect costs and the unit costs for the various utilities are identical to those used previously for the other potential feedstocks. Also, in a similar manner to that used for the other feedstocks, three prices for the heavy oil were assumed. The lowest cost, \$0.35/gal, is the current market price while the medium cost (\$0.53/gal) and the high cost (\$0.70/gal) options were included to cover higher price scenarios. The resulting unit production cost for NH_3 ranged from about \$217/ton for the low cost case (\$0.35/gal) to about \$302/ton for the high cost fuel-oil case (\$0.70/gal). The detailed breakdown of the annual operating costs for the low cost heavy fuel-oil case is shown in Table 37. The breakdowns for the other two feedstock costs are shown in Tables 38 and 39.

For the \$0.35/gal case, the major annual operating cost (39.2%) was for heavy fuel oil. Capital charges required 36.8% of the total annual amount. The other annual charges each represent less than 8% of the total operating costs.

NH_3 FROM COAL (47, 52, 53)

With the recent concern about the availability and cost of natural gas in the United States, increasing attention has been focused on the potential use of coal as an H_2 feedstock for the production of NH_3. Before the discovery of large natural gas deposits in the early 1950's, the primary feedstock for NH_3 was coal.

The major difference between using coal and the current use of natural gas would be the replacement of the primary and secondary reformers with coal gasifiers, coal preparation equipment, and crude gas cleanup system, as shown in Figure 16. The major advantage of using coal gasification would be an ample and long-term domestic supply of feedstock. The major disadvantage would be the increased capital investment for the gasifiers, the coal preparation equipment, and the gas purification system.

Although the following sections will discuss the commercially proven NH_3-from-coal processes, several demonstration projects involving new coal gasification technology are in various stages of development. The demonstration at TVA's National Fertilizer Development Center of retrofitting a coal gasification system onto the front end of an existing natural gas-based NH_3 plant will be the first of these modern NH_3-from-coal facilities to be operating in the United States. Initial operation of this plant, which will convert 168 tons/day of coal to NH_3, is scheduled for early 1980. Another project is a grass roots 1200 ton/day NH_3 plant consuming 1700 tons/day of coal. The contract for this plant between the U.S. Department of Energy (DOE) and W. R. Grace and Company, in conjunction with Ebasco Services, Inc., involves a feasibility study for the coal gasification and gas purification sections of the NH_3 plant. If the results of the feasibility study are favorable and this project is selected for further funding, operation of the plant could begin as early as 1982.

Process Description

Coal gasification is essentially a partial-oxidation operation similar to that used in the treatment of heavy fuel oil feedstock. Pulverized coal, steam, and O_2 from an air separation unit are fed to a gasifier, which can be a fixed-bed, an entrained-bed, or a fluidized-bed reactor. Although many

TABLE 37. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING HEAVY FUEL OIL AS THE H_2 FEEDSTOCK[a]

(LOW FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Heavy fuel oil	6,715,000 MBtu	2.43/MBtu	16,317,000	22.8
Catalyst and chemicals			150,000	0.2
Total raw materials costs			16,467,000	23.0
Conversion				
Operating labor and supervision	74,800 man-hr	10.00/man-hr	748,000	1.0
Utilities				
Heavy fuel oil	4,820,000 MBtu	2.43/MBtu	11,713,000	16.4
Electricity	45,000,000 kWh	0.027/kWh	1,215,000	1.7
Cooling water	26,400,000 kgal	0.11/kgal	2,904,000	4.1
Boiler feedwater	120,000 kgal	0.92/kgal	110,000	0.2
Maintenance, 5% of plant investment			5,680,000	7.9
Analyses	2,080 man-hr	15.00/man-hr	31,000	–[b]
Total conversion costs			22,401,000	31.3
Total direct costs			38,868,000	54.3
Indirect Costs				
Capital charges, 22.1% of capital investment			26,365,000	36.8
Overhead				
Plant, 50% of conversion costs less utilities			3,230,000	4.5
Administrative, 15% of operating labor and maintenance			969,000	1.4
Marketing			731,000	1.0
Research and development, 2% of direct and indirect costs			1,432,000	2.0
Total indirect costs			32,727,000	45.7
Total annual operating costs			71,595,000	100.0
	$/ton of NH_3			
Equivalent unit operating costs	216.95			

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a. Basis (47)
Total capital investment, $119,300,000.
Total plant investment, $113,600,000.
On-stream time, 330 days/yr.

b. Less than 0.1%.

TABLE 38. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING HEAVY FUEL OIL AS THE H_2 FEEDSTOCK[a]

(MEDIUM FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Heavy fuel oil	6,715,000 MBtu	3.65/MBtu	24,510,000	28.6
Catalyst and chemicals			150,000	0.2
Total raw materials costs			24,660,000	28.8
Conversion				
Operating labor and supervision	74,800 man-hr	10.00/man-hr	748,000	0.9
Utilities				
Heavy fuel oil	4,820,000 MBtu	3.65/MBtu	17,593,000	20.6
Electricity	45,000,000 kWh	0.027/kWh	1,215,000	1.4
Cooling water	26,400,000 kgal	0.11/kgal	2,904,000	3.4
Boiler feedwater	120,000 kgal	0.92/kgal	110,000	0.1
Maintenance, 5% of plant investment			5,680,000	6.6
Analyses	2,080 man-hr	15.00/man-hr	31,000	_[b]
Total conversion costs			28,281,000	33.0
Total direct costs			52,941,000	61.8
Indirect Costs				
Capital charges, 22.1% of capital investment			26,365,000	30.7
Overhead				
Plant, 50% of conversion costs less utilities			3,230,000	3.8
Administrative, 15% of operating labor and maintenance			969,000	1.1
Marketing			731,000	0.9
Research and development			1,432,000	1.7
Total indirect costs			32,727,000	38.2
Total annual operating costs			85,668,000	100.0
Equivalent unit operating costs	$/ton of NH_3 259.60			

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[a] Basis (47)
Total capital investment, $119,300,000.
Total plant investment, $113,600,000.
On-stream time, 330 days/yr.

[b] Less than 0.1%.

TABLE 39. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING HEAVY FUEL OIL AS THE H_2 FEEDSTOCK[a]

(HIGH FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Heavy fuel oil	6,715,000 MBtu	4.86/MBtu	32,635,000	32.7
Catalyst and chemicals			150,000	0.2
Total raw materials costs			32,785,000	32.9
Conversion				
Operating labor and supervision	74,800 man-hr	10.00/man-hr	748,000	0.8
Utilities				
Heavy fuel oil	4,820,000 MBtu	4.86/MBtu	23,425,000	23.5
Electricity	45,000,000 kWh	0.027/kWh	1,215,000	1.2
Cooling water	26,400,000 kgal	0.11/kgal	2,904,000	2.9
Boiler feedwater	120,000 kgal	0.92/kgal	110,000	0.1
Maintenance, 5% of plant investment			5,680,000	5.7
Analyses	2,080 man-hr	15.00/man-hr	31,000	–[b]
Total conversion costs			34,113,000	34.2
Total direct costs			66,898,000	67.1
Indirect Costs				
Capital charges, 22.1% of capital investment			26,365,000	26.5
Overhead				
Plant, 50% of conversion costs less utilities			3,230,000	3.3
Administrative, 15% of operating labor and maintenance			969,000	1.0
Marketing			731,000	0.7
Research and development			1,432,000	1.4
Total indirect costs			32,727,000	32.9
Total annual operating costs			99,625,000	100.0
Equivalent unit operating costs		\$/ton of NH_3 301.89		

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a. Basis (47)
Total capital investment, \$119,300,000.
Total plant investment, \$113,600,000.
On-stream time, 330 days/yr.

b. Less than 0.1%.

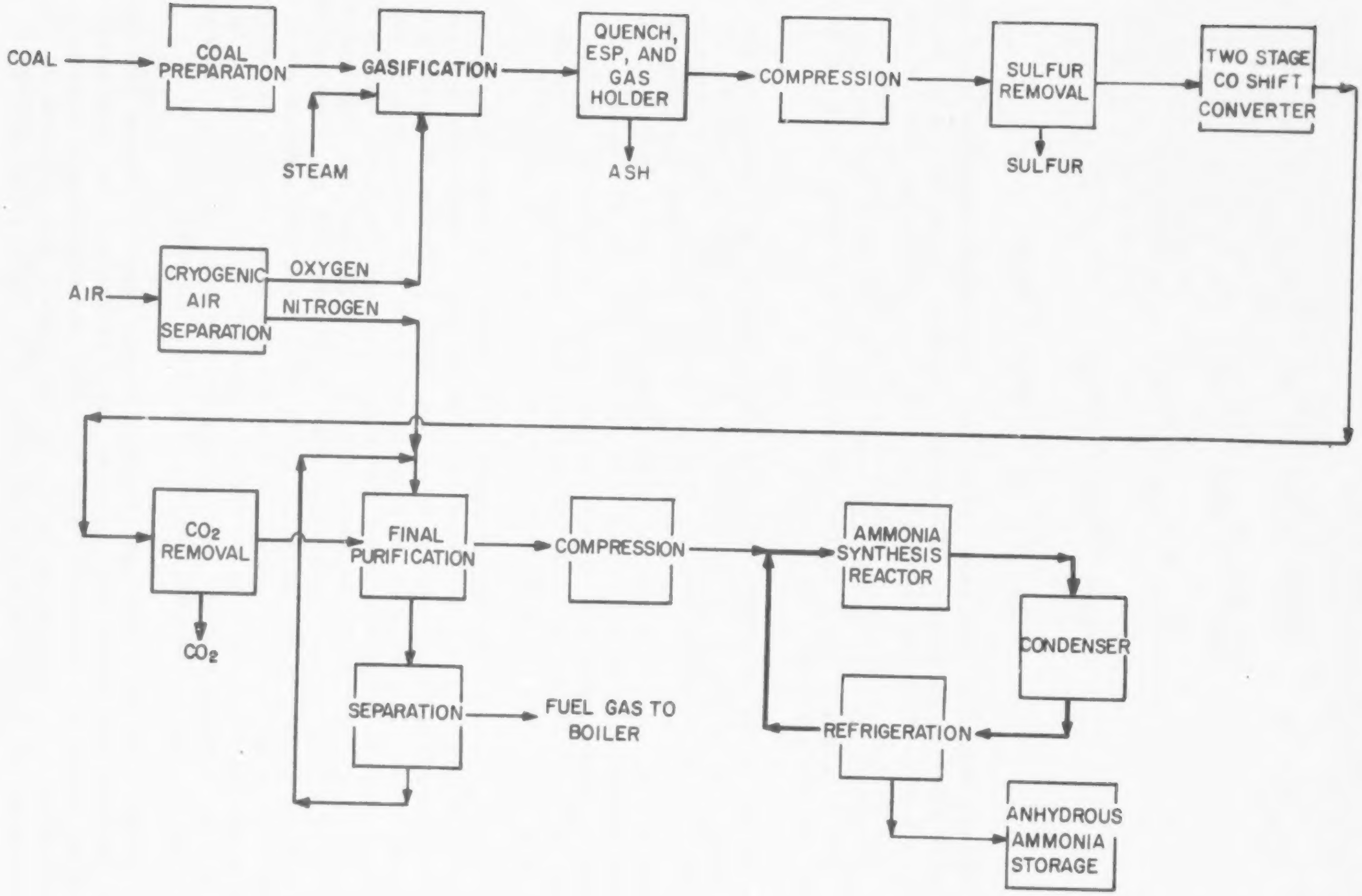

Figure 16. Block flow diagram of a typical NH_3 plant based on coal as the H_2 feedstock.

chemical reactions involving coal, steam, and O_2 occur in the gasifier, the primary equilibrium reactions are:

$$C_{(s)} + H_2O_{(g)} \rightleftarrows CO_{(g)} + H_{2(g)} \quad (10)$$

$$C_{(s)} + 1/2O_{2(g)} \rightleftarrows CO_{(g)} \quad (11)$$

$$C_{(s)} + O_{2(g)} \rightleftarrows CO_{2(g)} \quad (12)$$

$$C_{(s)} + CO_{2(g)} \rightleftarrows 2CO_{(g)} \quad (13)$$

$$CO_{(g)} + H_2O_{(g)} \rightleftarrows CO_{2(g)} + H_{2(g)} \quad (14)$$

The desired reaction, reaction (10), is an endothermic reaction which requires extensive amounts of energy. The primary reason for injecting O_2 is to generate sufficient heat from reactions (11) and (12) to allow reaction (10) to proceed. The final gaseous mixture, or crude gas, is a complex equilibrium system that is dependent on the degree of oxidation, the temperature and pressure, and hence the particular gasification system selected as well as the feed ratios of the various streams.

The gasification scheme selected has a significant impact on the various downstream operations prior to entering the NH_3 synthesis loop and therefore two gasification variations will be briefly mentioned. The first is a fixed-bed reactor, such as the Lurgi gasifier. These gasifiers can operate at low or high pressure but are typically low-temperature units (800-1100°C). The second variation is an entrained flow reactor such as the Koppers-Totzek gasifier. These gasifiers operate at low pressure at a high temperature (1600-2000°C). Other gasification systems and additional data on the above-mentioned gasifiers are available from other sources (11, 35, 37).

Although generating NH_3 synthesis gas using low-temperature gasifiers is technically feasible and has been used in the past, these type systems may not be attractive for producing NH_3 since these processes generate more of the higher hydrocarbons such as methane and tars and hence less H_2 and CO. The higher operating temperatures of the entrained-flow reactors, on the other hand, convert the coal to primarily H_2 and CO with very few higher hydrocarbons. Typical exit gas compositions after initial cleanup for the low- and high-temperature gasifiers are shown in Table 40. These differences would be substantially larger, of course, in the raw gas since the low-temperature gasifier product gas also contains tars, phenols, and NH_3 which are removed in the initial gas cleanup section. For these reasons only the high-temperature gasification system (Koppers-Totzek) will be discussed further.

In the Koppers-Totzek gasification scheme shown in Figure 16, the pulverized coal, oxygen, and stream are fed into the gasifier and the previously mentioned reactions occur at 1600-2000°C (2912-3632°F) and atmospheric pressure. Significant amounts of ash are entrained in the product gas and must be removed. This product is initially cooled to about 870-1100°C (1598-2012°F) by a water quench and then passed through a waste heat boiler to further cool the gas to 300°C (572°F). Final cooling to about 40°C (104°F) is achieved in a cooler washer and the resulting gas is sent to the ESP for additional particulate removal. This partially cleaned gas passes through a gas holder and is then compressed to about 30 atmospheres prior to chemical purification.

TABLE 40. TYPICAL EXIT GAS COMPOSITIONS FOR

LOW AND HIGH TEMPERATURE GASIFIERS (36, 53)

	Composition, vol %[a]	
Constituent	Low temperature	High temperature
CO_2	30.3	13.4
CO	17.5	54.4
H_2	42.1	30.0
CH_4	9.6	0.1
H_2S + COS	-[b]	0.6
N_2 + Ar	0.5	1.5

a. Dry basis.

b. Not listed.

Since coal contains significant amounts of sulfur which will be present in the synthesis gas primarily as H_2S, but with lesser amounts of COS, CS_2, and other miscellaneous carbon-sulfur compounds, these compounds are removed prior to shift conversion. Although many H_2S removal systems are commercially available (including the hot K_2CO3 process described earlier), the Rectisol processes will be described here. In the Rectisol process, the H_2S-containing stream is passed countercurrently to a cold (-30 to -50°C) methanol solution which preferentially and almost completely removes H_2S. The resulting H_2S-containing methanol stream also contains some CO_2 and can be regenerated in a separate vessel by reducing the pressure. This concentrated H_2S stream can then be further processed to form byproduct elemental sulfur in a Claus unit while the methanol is recycled to the absorber. The "partially" cleaned synthesis gas, with the typical composition shown in Table 41, is then preheated, injected with steam, and sent to a conventional two-stage CO shift converter.

The first-stage converter uses an iron-chromium (Fe-Cr) catalyst at 415° (750°) to promote the water gas shift reaction, i.e.,

$$CO_{(g)} + H_2O_{(g)} \rightarrow CO_{2(g)} + H_{2(g)} \qquad (15)$$

and thereby producing as much H_2 as possible while at the same time consuming CO. After cooling to about 250°C (482°F) in an interstage heat exchanger, the crude gas is passed through a low-temperature catalytic reactor containing a zinc-copper oxide (ZnO-CuO) catalyst to further reduce the final CO concentration. In addition to the interstage cooling, the crude gas is usually passed through a guard cell containing ZnO for Fe_2O_3 to remove the last traces of sulfur compounds before entering the low-temperature shift converter. The typical composition of this shifted gas is given in Table 41.

TABLE 41. TYPICAL SYNTHESIS GAS COMPOSITION

BEFORE AND AFTER CO SHIFT (53)

	Composition, vol %[a]	
Constituent	Before shift conversion	After shift conversion
CO_2	11.3	41.3
CO	56.0	3.0
H_2	31.0	54.6
N_2	1.0	0.6
Ar	0.6	0.4
CH_4	0.1	0.1

a. Dry basis.

The synthesis gas is then passed through a final purification stage consisting of CO_2 removal system and a cryogenic N_2 wash. This CO_2 removal system is another low-temperature (-54°C) methanol wash nearly identical with that used for H_2S removal. The cold methanol containing CO_2 is pumped to a second vessel where, at a reduced pressure, the CO_2 is flashed off as a byproduct stream. The regenerated methanol is then recycled to the absorber. The synthesis gas leaving the CO_2 wash, which has a typical composition, shown in Table 42, passes through carbon beds to remove any remaining CO_2 and methanol which would freeze out at cryogenic temperatures.

As a final purification stage and also to provide the N_2 for the 3:1 stoichiometric ratio of N_2:H_2 for NH_3 production, the gas is sent to a liquid N_2 washing. This final purification stage is identical with that previously discussed in the section on partial oxidation. The resulting synthesis gas with the composition shown in Table 42 is further compressed before entering the conventional NH_3 synthesis loop. A portion (13-17%) of the synthesis mixture is converted per pass through the reactor with the remainder recycled for the addition of more synthesis gas mixture and another pass through the reactor. The NH_3 is separated by condensation and stored as a liquid in refrigerated storage tanks.

TABLE 42. TYPICAL SYNTHESIS GAS COMPOSITIONS

AFTER CO_2 WASH AND AFTER LIQUID N_2 WASH (59)

	Composition, vol %[a]	
Constituent	After CO_2 wash	After liquid N_2 wash
CO_2	-	-
CO	5.0	-
H_2	93.1	75
N_2	1.1	25
Ar	0.6	-
CH_4	0.1	-

a. Dry basis.

Recently Texaco Development Company has begun to market a coal gasifier which is gaining acceptance as an additional alternative to the Koppers-Totzek gasification system. The Texaco process, although a lower temperature process [gasifier temperature, 1100-1350°C (2012-2462°F)], uses a high pressure (500 psig) gasifier and thus eliminates some of the downstream compression stages. The raw gas from the gasifier has a composition similar to that generated in the high-temperature systems as shown in Table 43. The remaining processing steps are similar to those discussed for the high-temperature gasification process, i.e., desulfurization, CO shift, CO_2 removal, final purification, compression, and NH_3 synthesis.

TABLE 43. COMPARISON OF TYPICAL RAW GAS COMPOSITION

FROM THE TEXACO WITH THAT FOR A HIGH TEMPERATURE GASIFIER[a] (53)

	Gas compositions, vol %	
Constituent	Texaco	High temperature
CH_4	0.2	0.1
CO	45.0	54.4
CO_2	17.2	13.4
H_2	35.8	30.0
N_2 + Ar	0.6	1.5
H_2S + COS	1.2	0.6

a. Dry basis.

Raw Material, Utility, Labor, and Energy Requirements

The raw material, utility, labor, and energy requirements for producing 1 ton of NH_3 in a 1000 ton/day NH_3-from-coal plant are given in Table 44. As is typical for NH_3-from-coal plants, the primary raw material and utility requirement is coal with this coal also accounting for essentially all (98.9%) of the input energy.

TABLE 44. UNIT RAW MATERIAL, UTILITY, LABOR, AND ENERGY

REQUIREMENTS FOR THE PRODUCTION OF NH_3 FROM COAL[a] (47)

Item	Quantity required, unit/ton NH_3	Energy equivalent, MBtu/ton NH_3
Raw material		
Coal, MBtu	27.4	27.4
Utility		
Coal, MBtu	13.7	13.7
Electricity, kWh	136.0	0.5[b]
Cooling water, kgal	70.0	-
Boiler feedwater, kgal	0.45	-
Labor, man-hr	0.45	-
Total		41.6

a. Basis
Plant capacity, 1000 tons/day.
On-stream time, 330 days/yr.
b. Assuming 3,412 Btu/kWh.

Process Economics

The total capital investment for a 1000 ton/day NH_3-from-coal plant was estimated at $132.8 M based on information given by others (47). As in all the previous economic summaries, the annual operating costs were calculated on the basis of joint EPA-TVA premises and 1978 raw material and utility costs. If a coal cost of $20.00/ton is assumed, the resulting annual operating costs are approximately $61.4 M and the equivalent unit operating cost is about $186/ton of NH_3 as shown in Table 45. Capital charges represented about 47.8% of the annual operating cost while coal cost contributed about 21%. The remaining annual costs each represented 11% or less of the total annual cost.

TABLE 45. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING COAL AS THE H_2 FEEDSTOCK[a]

(LOW FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Coal	9,029,000 MBtu	0.95/MBtu	8,578,000	14.0
Catalyst and chemicals			150,000	0.2
Total raw materials costs			8,728,000	14.2
Conversion				
Operating labor and supervision	150,000 man-hr	10.00/man-hr	1,500,000	2.4
Utilities				
Coal	4,514,000 MBtu	0.95/MBtu	4,288,000	7.0
Electricity	45,000,000 kWh	0.027/kWh	1,215,000	2.0
Cooling water	23,100,000 kgal	0.11/kgal	2,541,000	4.2
Boiler feedwater	150,000 kgal	0.92/kgal	138,000	0.2
Maintenance, 5% of plant investment			6,430,000	10.5
Analyses	2,080 man-hr	15.00/man-hr	31,000	_[b]
Total conversion costs			16,143,000	26.3
Total direct costs			24,871,000	40.5
Indirect Costs				
Capital charges, 22.1% of capital investment			29.349,000	47.8
Overhead				
Plant, 50% of conversion costs less utilities			3,981,000	6.5
Administrative, 15% of operating labor and maintenance			1,194,000	2.0
Marketing			731,000	1.2
Research and development, 2% of direct and indirect costs			1,227,000	2.0
Total indirect costs			36,482,000	59.5
Total annual operating costs			61,353,000	100.0
	$/ton of NH_3			
Equivalent unit operating costs	185.92			

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a. Basis (47)
Total capital investment, $132,800,000.
Total plant investment, $128,600,000.
On-stream time, 330 days/yr.

b. Less than 0.1%.

Alternative coal cost of \$30/ton and \$40/ton were chosen to reflect potentially higher fuel costs and the resulting annual operating cost for each case is shown in Table 46 and 47 respectively.

NH_3 IMPORTS (7, 13, 14)

In addition to domestic NH_3 production, net imports or exports have an additional impact on the U.S. NH3 supply and demand situation. Until 1973, the United States was a perennial net exporter of NH_3 as shown in Figure 17. The high levels of U.S. exports during the period 1967-1971 correspond to the time when many of the new large capacity (1000 ton/day) NH_3 plants were coming onstream and there was a large surplus of domestic capacity. With this surplus of NH_3 capacity no new plants were initiated; as demand increased to absorb this excess capacity, the United States became a net importer of NH_3 from 1975-1977. The recent large increase in domestic NH_3 is expected to sharply curtail NH_3 imports during 1978 and 1979, and since no new plants are currently under construction, imports of NH3 are projected to rise sharply again in the 1980's.

Most of these NH_3 imports are expected to come from three countries: Russia, Canada, and Mexico. Of a projected total 3.5 Mtons of NH_3 imports (gross) in 1980, an estimated 1.5 Mtons will come from Russia, 0.5 Mtons from Canada, and 1.5 Mtons from Mexico and other Latin American companies. The Russian imports are the result of long-term contract for a phosphate-N_2 trade negotiated between Occidental Petroleum and Russia. The Canadiam imports of 0.5 Mtons in 1980 is about double the current rate and is primarily due to the large increases in installed NH_3 capacity expected to come online during the next 2 yr. However, most of these Canadian imports are consumed across the border in the upper Midwest and will not contribute significantly to the potential NH_3 oversupply on the gulf coast. The Mexican imports, however, are perhaps the most unsettling from the standpoint of the U.S. NH_3 producers. With its recently discovered large oil and gas reserves (total reserves of 20 Tft^3 in the Reforma field alone) and the projected large NH_3 plant capacity additions, not only will Mexico have the capacity to export large quantities of NH_3 to the U.S. Gulf and East Coasts (as well as the U.S. West Coast through a trans-Mexican NH_3 pipeline) but many of the export orders by the other Latin American countries will be filled with Mexican NH_3.

Projected Mexican NH_3 Capacity and Production

With their large and increasing natural gas production and reserves, Mexico has the potential to become an important exporter of NH_3 as well as natural gas. The proximity of their gas-producing areas to the Gulf of Mexico, and hence the ease of shipment to the U.S. Gulf Coast and to other Latin American countries by barge would seem to offer attractive economic incentive to PEMEX (Petroleos Mexicano's), particularly since they have recently begun an unprecedented growth in NH_3 capacity. Six new plants, listed in Table 48, are in various stages of construction and will raise the total annual NH_3 production in Mexico to nearly 4.0 Mtons.

TABLE 46. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING COAL AS THE H_2 FEEDSTOCK[a]

(MEDIUM FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Coal	9,029,000 MBtu	1.43/MBtu	12,911,000	19.0
Catalyst and chemicals			150,000	0.2
Total raw materials costs			13,061,000	19.2
Conversion				
Operating labor and supervision	150,000 man-hr	10.00/man-hr	1,500,000	2.2
Utilities				
Coal	4,514,000 MBtu	1.43/MBtu	6,455,000	9.5
Electricity	45,000,000 kWh	0.027/kWh	1,215,000	1.8
Cooling water	23,100,000 kgal	0.11/kgal	2,541,000	3.8
Boiler feedwater	150,000 kgal	0.92/kgal	138,000	0.2
Maintenance, 5% of plant investment			6,430,000	9.5
Analyses	2,080 man-hr	15.00/man-hr	31,000	–[b]
Total conversion costs			18,310,000	27.0
Total direct costs			31,371,000	46.2
Indirect Costs				
Capital charges, 22.1% of capital investment			29,349,000	43.2
Overhead				
Plant, 50% of conversion costs less utilities			3,981,000	5.9
Administrative, 15% of operating labor and maintenance			1,194,000	1.8
Marketing			731,000	1.1
Research and development			1,227,000	1.8
Total indirect costs			36,482,000	53.8
Total annual operating costs			67,853,000	100.0
Equivalent unit operating costs		\$/ton of NH_3 205.62		

a. Basis (47)
Total capital investment, \$132,800,000.
Total plant investment, \$128,600,000.
On-stream time, 330 days/yr.
b. Less than 0.1%.

TABLE 47. TOTAL ANNUAL OPERATING COSTS AND UNIT OPERATING COSTS FOR A 1000 TON/DAY NH_3 PLANT USING COAL AS THE H_2 FEEDSTOCK[a]

(HIGH FEEDSTOCK COST CASE)

Item	Annual quantity (47)	Unit cost, $	Total annual operating cost, $	Percent of total cost
Direct Costs				
Raw materials				
Coal	9,029,000 MBtu	1.90/MBtu	17,155,000	23.1
Catalyst and chemicals			150,000	0.2
Total raw materials costs			17,305,000	23.3
Conversion				
Operating labor and supervision	150,000 man-hr	10.00/man-hr	1,500,000	2.0
Utilities				
Coal	4,514,000 MBtu	1.90/MBtu	8,577,000	11.6
Electricity	45,000,000 kWh	0.027/kWh	1,215,000	1.6
Cooling water	23,100,000 kgal	0.11/kgal	2,541,000	3.4
Boiler feedwater	150,000 kgal	0.92/kgal	138,000	0.2
Maintenance, 5% of plant investment			6,430,000	8.7
Analyses	2,080 man-hr	15.00/man-hr	31,000	–[b]
Total conversion costs			20,432,000	27.5
Total direct costs			37,737,000	50.8
Indirect Costs				
Capital charges, 22.1% of capital investment			29,349,000	39.5
Overhead				
Plant, 50% of conversion costs less utilities			3,981,000	5.4
Administrative, 15% of operating labor and maintenance			1,194,000	1.6
Marketing			731,000	1.0
Research and development			1,227,000	1.7
Total indirect costs			36,482,000	49.2
Total annual operating costs			74,219,000	100.0
		$/ton of NH_3		
Equivalent unit operating costs		224.91		

a. Basis (47)
Total capital investment, $132,800,000.
Total plant investment, $128,600,000.
On-stream time, 330 days/yr.

b. Less than 0.1%.

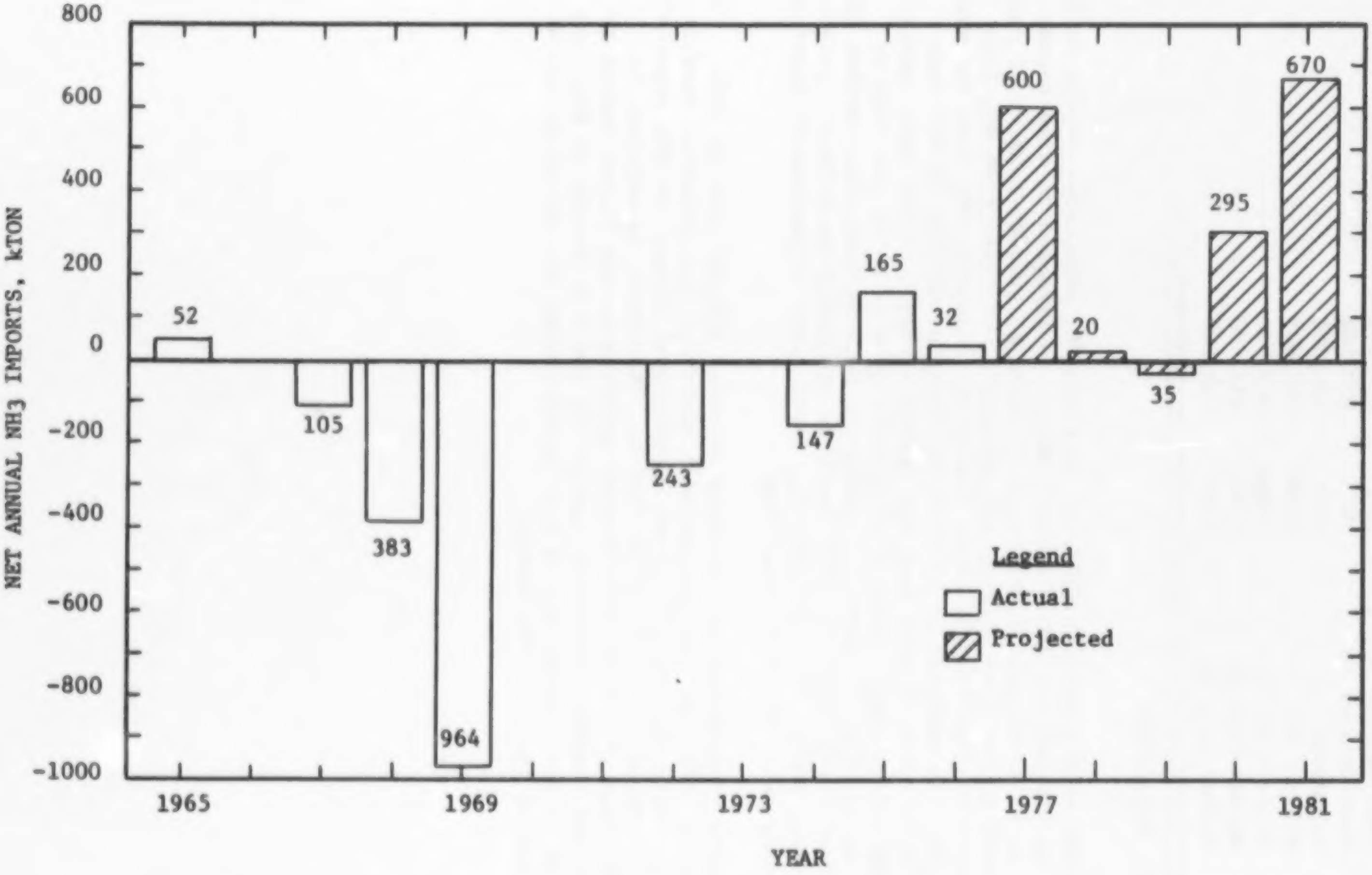

Figure 17. Net U.S. imports and exports of NH_3 during the period 1965-1981 (43).

TABLE 48. PROJECTED MEXICAN NH_3 PLANTS (8, 19)

Plant	Startup date	Capacity, tons/yr
Cosoleacaque No. 1	1977	450,000
Cosoleacaque No. 2	1977	450,000
Salamanca	1978	300,000
Allende No. 1	1980[a]	445,000
Allende No. 2	1980[a]	445,000
Salima Cruz No. 1	1980[a]	455,000
Salima Cruz No. 2	1980[a]	445,000

a. Projected.

Even though only a portion (40%) of this new NH_3 production coming online in 1977 will be comsumed to produce urea, Mexico is expected to also become a net exporter of this valuable fertilizer. A total of 460,000 tons/yr of this NH_3 will be used in two urea plants, one using 275,000 tons/yr of NH_3 from the Cosoleacaque complex and the other urea plant consuming NH_3 from the new Salamanca plant. In contrast to 1975-76 when Mexico imported 46,000 tons of nitrogen fertilizers (mostly from the United States), in the near future, Mexico is expected to export upwards of 200,000 tons of N_2 in the form of urea. Thus, in addition to possible losing the Mexican fertilizer market (50,000 tons of nitrogen fertilizers in 1976) American producers can expect a potential double loss since Mexico will be exporting fertilizers to previously American markets in other Latin American countries.

NH_3 exports from Mexico are expected to reach 500,000 tons by 1978. For the years beyond 1978 numerous factors will determine the potential Mexican exports of NH_3 including population and agricultural trends but NH_3 exports could reach 1.5 Mtons/yr in the early 1980's. Therefore, in addition to displacing NH_3 (and NH_3-based fertilizers) exported by the United States to South America and thereby indirectly increasing the U.S. supply of NH_3, the export of some of this excess NH_3 to the United States has the potential of directly affecting the U.S. NH_3 supply.

ECONOMIC COMPARISON OF NH_3 PRODUCTION COST FOR VARIOUS H_2 FEEDSTOCKS

From a comparison of the estimated 1978 NH_3 production costs for the various hydrocarbon feedstocks calculated in the previous sections and summarized in Table 49, it is apparent that these alternate feedstocks cannot economically compete with natural gas at the present time based on the premises of this study. With coal priced at \$20.00/ton (0.95/MBtu) and naphtha costing \$0.40/gal (\$3.00/MBtu), the price of natural gas must exceed about \$3.40/MBtu in order for coal and \$3.60 for naphtha to become economically attractive.

Although the price of natural gas is expected to increase in the future, the actual selling price of natural gas is uncertain and hence the relative price difference between natural gas and the other alternate feedstocks are difficult to project into the future. Even the future availability of natural gas for NH_3 production is questionable. The economics of NH_3 production from the various feedstocks in the future will depend not only on the availability and cost of feedstocks, but also on innovations resulting from advanced process development.

TABLE 49. ALTERNATIVE FEEDSTOCK COSTS AND THE RESULTING UNIT PRODUCTION COST OF NH_3 IN 1978

Feedstock	Assumed cost $/unit	 $/MBtu	Unit production cost of NH_3,[a] $/ton
Natural gas	1.00/MBtu	1.00	111
	3.00/MBtu	3.00	174
	5.00/MBtu	5.00	237
Naphtha	0.40/gal	3.00	193
	0.60/gal	4.50	244
	0.80/gal	6.00	294
Fuel oil	0.35/gal	2.43	217
	0.53/gal	3.65	260
	0.70/gal	4.86	302
Coal	20.00/ton	0.95	186
	30.00/ton	1.43	206
	40.00/ton	1.90	225

[a] Basis

Plant capacity, 1000 tons/day.
On-stream time, 330 days/yr.
Premises are described in text under "Current Generating Techniques," "Steam Reforming of Natural Gas."

CURRENT AND PROJECTED NATURAL GAS SUPPLY AND DEMAND

At one time in the United States (and even at the present time in some countries) natural gas was vented or flared because there were no local markets for the gas but the associated oil was highly desired. With the increase in drilling for oil in the 1950's in the United States and the associated increase in natural gas supplies, transcontinental pipelines were constructed so that the resulting gas byproduct could be transported to areas of the country without substantial natural gas reserves. Since natural gas was orginally considered as a waste stream from oil wells, this gas was sold to the pipeline companies at low prices simply for disposal. The demand for natural gas continued to grow as more and more consumers switched from coal and coal-gas to the cheaper and cleaner natural gas. This change over was significantly increased during the rise of the environmental movement of the early 1970's.

With the major multinational oil companies concentrating on drilling and production the total number of wells being drilled in the United States gradually dropped during the 1960's and early 1970's. The proven natural gas reserves began to decline as more gas was consumed than was being replaced by the discovery of new natural gas reservoirs. Retail gas companies continued to sign up both residential and industrial customers during this time until natural gas demand began to exceed pipeline supplies during the winter months and industrial customers faced substantial curtailments.

This trend has led to the current situation--industrial customers consuming natural gas from interstate pipelines face substantial natural gas cutbacks during the winter months when residential demands require most of the available gas. This winter curtailment is a substantial operating problem for natural gas-based NH_3 plants since the peak demand and sales period for fertilizers is during spring planting and it would be desirable for the NH_3 plants to be operating at peak capacity during these winter months.

One alternative solution to this natural gas shortage problem which has already been taken by many of the NH_3 producers is the substitution of intrastate for interstate natural gas. NH_3 plants operating prior to 1976, i.e., NH_3 plants planned and designed before the Arab oil embargo and the era of natural gas curtailments, were scattered around the country but were concentrated primarily in the Gulf States (gas-producing states) or states consuming large amounts of NH_3, i.e., the Corn Belt States such as Nebraska and Iowa, as shown in Figure 18. Figure 19 pinpoints the location of NH_3 plants coming online during the period 1976-1980, i.e., plants planned after the oil embargo and the beginning of the winter gas curtailments. At the time these NH_3 plants were being planned interstate gas, although slightly cheaper, was subject to these winter curtailments. The pipeline gas companies selling interstate gas were banned from accepting new customers and the new post-1975 NH_3 plants were built primarily in gas-producing states such as Oklahoma, Louisiana, and Texas. However,

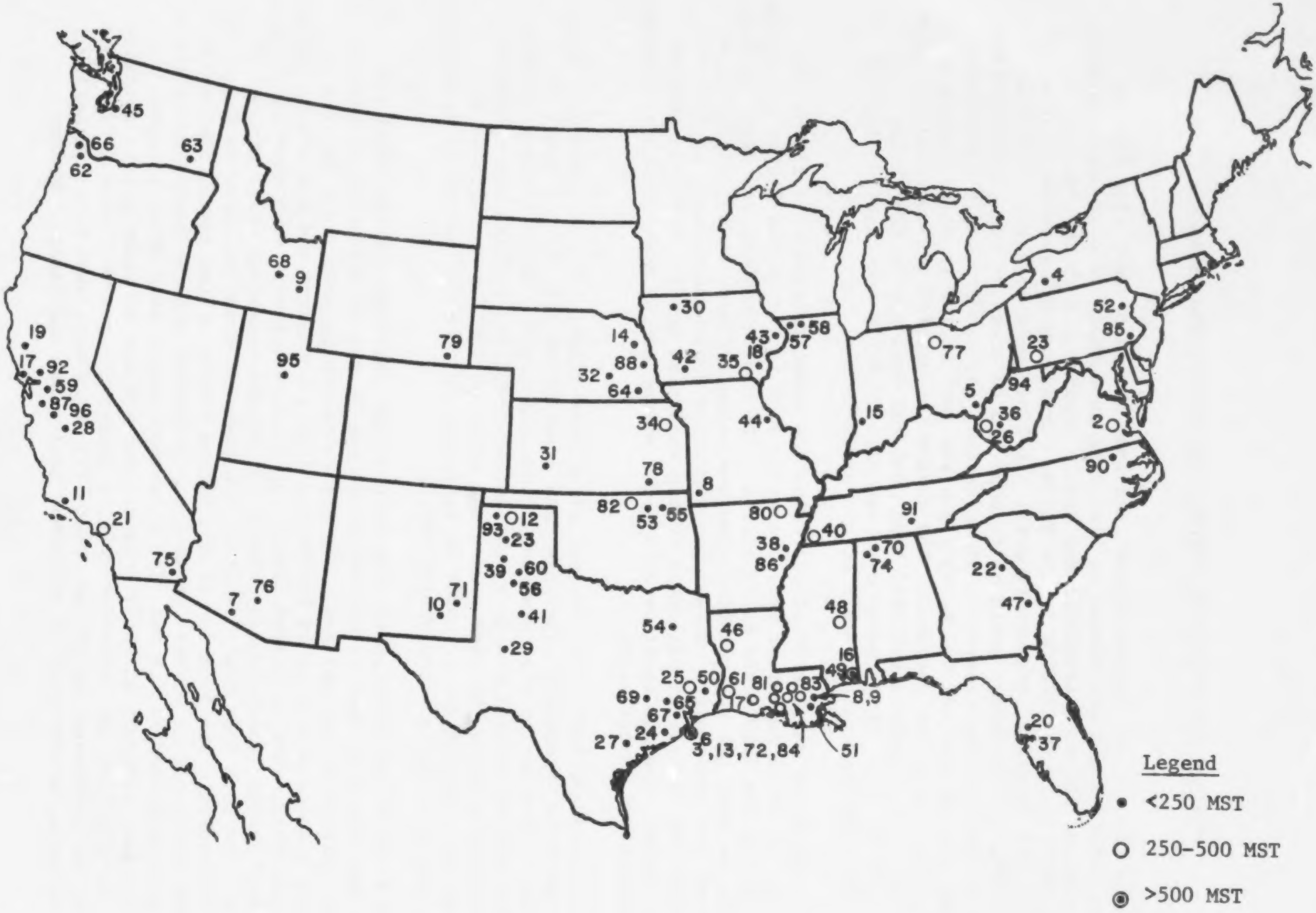

Figure 18. Geographic location of U.S. NH_3 plants operating before 1975 (39).

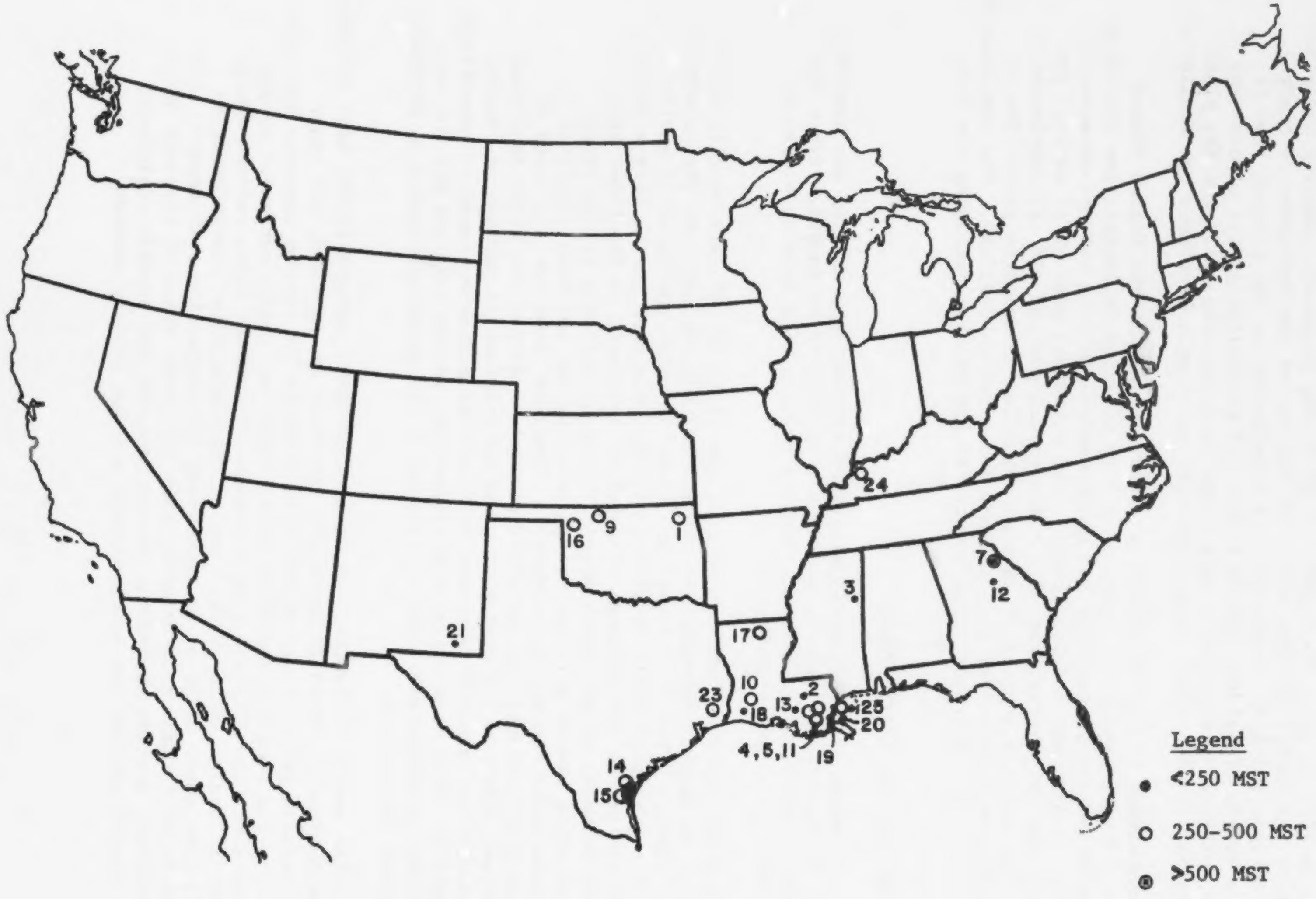

Figure 19. Geographic location of new U.S. NH_3 plants operating after 1975 (39).

with the recent substantial increases in intrastate gas prices (more than \$2.00/MBtu in 1977 vs about \$0.55/MBtu in 1973) NH_3 producers planning new plants may return to the interstate gas market if the ban on new customers is removed. The other options which NH_3 producers are considering at the present time to guarantee a supply of natural gas include: (1) diversifying into natural gas production (or form a partnership with a gas producer) and build the NH_3 plant or (2) locating the NH_3 plant where long-term gas supply contracts are available.

Any assessment of the availability of natural gas in the United States during the period 1980-2000 in essence will be affected by legislative decisions concerning future energy supplies and environmental factors and not necessarily on technical or economic considerations. The potential sources of natural gas are divided into two categories: (1) conventional sources and (2) supplemental sources. Conventional sources of natural gas refer to domestic production in the contiguous 48 states from both onshore and offshore drilling. The supplemental sources include all other potential sources of natural gas including gas from Alaska.

CONVENTIONAL NATURAL GAS SUPPLIES

The assessment of future availability of conventional natural gas depends, of course, on the actual reservoirs of gas available to be tapped. Since the total domestic reserves are unknown, various methods have been developed to estimate these total gas reserves.

Hydrocarbon reserves, whether natural gas or crude oil, are usually split into several categories depending on the information available on the potential production field. "Proven" reserves are those in which enough wells and/or exploration has been carried out that producing this much gas is only a matter of pumping the gas to the surface. "Probable" reserves, on the other hand, incorporate reserves of gas which appear likely by drilling in the fringe areas of existing fields although enough drilling has not been completed to include these under the proven category. "Possible" reserves are based on information about potentially lucrative geological formations which have been identified and which in the past have provided substantial amounts of natural gas. "Speculative" reserves are more or less simple guesses based on essentially no information. Thus from a planning standpoint, proven reserves are the most important with progressively less confidence in the reserves classed as probable, possible, and speculative.

The total amount of natural gas in each of these categories has been estimated by various authors as shown in Table 50. As would be expected, the range of estimated natural gas reserves varies only slightly for proven reserves but quite substantially for possible and speculative reserves. These estimated proven reserves of about 216 Tft^3 are also a reason for concern since current annual production if running at 19-20 Tft^3 and at this rate these proven reserves would only last about 11 yr at current rates of consumption. Of course, this projection fails to take into consideration the other types of reserves which will undoubtedly provide additional proven reserves and provide additional years of natural gas but the exact amounts are now largely unknown.

TABLE 50. ESTIMATES OF U.S. NATURAL GAS RESERVES (5, 20, 31, 37, 40)

	Natural gas reserves, Tft^3, as estimated by:				
Type of reserve	FEA	Exxon	USGS	CIA	Potential gas committee
Proven	237	237	-	218	216
Probable	202	111	-	-	215
Possible	522	287	484	-	363
Speculative	-	295	-	-	345-395
Total	961	930	-	-	1,139-1,189

Further aggravating this concern about natural gas is the steady decline in proven reserves as shown in Figure 20 for the period 1968-1980. The amount of natural gas being produced and sold annually from these proven reserves is more than that being confirmed and added. Currently additional drilling is being done primarily in relatively safe areas, such as already existing fields where the success rate is high, and very little drilling is being done in previously unexplored areas (containing the possible or speculative reserves, most of the potential reserves listed in Table 50) where the risks for finding and proving large new gas fields are greater. Thus, under the present conditions the probability of substantially increasing proven reserves and thus allaying fears of a looming natural gas crisis are highly unlikely.

SUPPLEMENTAL NATURAL GAS SUPPLIES

In addition to these conventional natural gas deposits, other potential deposits of gas are domestically available but at substantially higher prices. These sources would include: (1) synthetic natural gas (SNG) from coal or oil, (2) imported natural gas from Alaska, Canada, and Mexico by pipeline or liquefied natural gas (LNG) from Algeria and Indonesia, and (3) new technologies such as gas from geopressurized brine solutions, natural gas from "tight" formations such as shales, and methane from coal mines. The major drawback associated with these alternative sources are the costs involved. The price of the resulting natural gas from these sources is expected to be greater than about \$3.00/kft^3 (vs recently increased rate for new interstate gas of \$1.44).

SNG From Oil and Coal

SNG from either oil or coal faces one other common problem in addition to the cost of the resulting gas. From the standpoint of making NH_3, it makes little sense to partially oxidize either feedstock to H_2 and CO, shift the gas mixture to methane, transport the gas through a pipeline system, and then steam reform the methane back into H_2 and CO. If either oil or coal is to be used to manufacture NH_3, from an overall economic and also technical point of view the

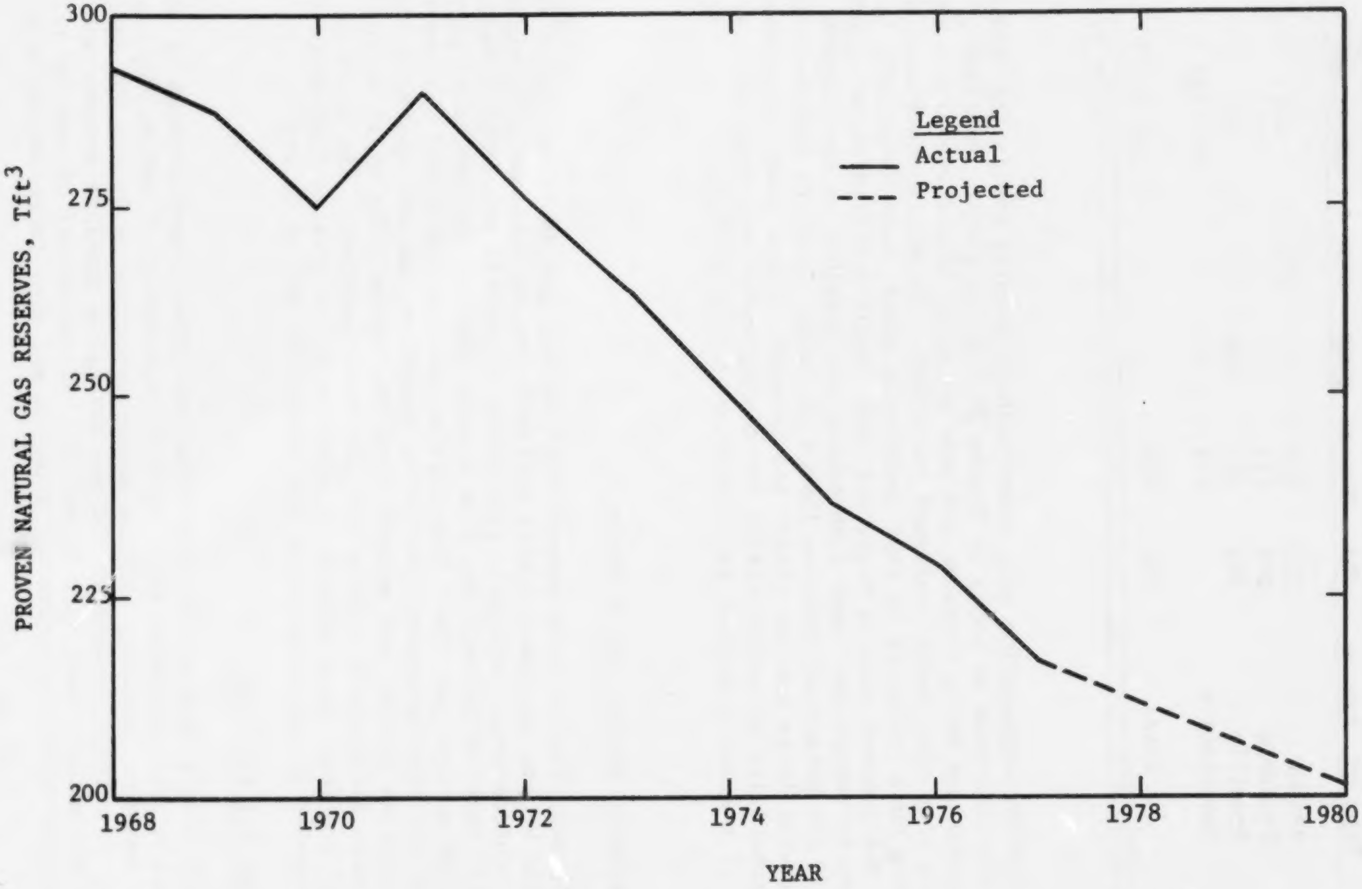

Figure 20. Proven U.S. natural gas reserves during the period 1968-1980 (51).

hydrocarbon feedstock should be gasified as the initial stage of the NH_3 unit and all of the resulting gas used directly (see earlier sections on partial oxidation of oil and coal) to make NH_3.

Another major problem associated with the use of petroleum fractions for making SNG is its desirability as a feedstock for other petrochemicals. The typical petroleum fraction used in generating SNG, natural gas liquids and naphtha, have many other more profitable uses. Natural gas liquids are a major feedstock for various plastic materials, as well as gasoline, while naphtha is used both in the blending gasoline and as a feedstock for various reforming operations. With the particular cut of naphtha used to make SNG already in short supply, the diversion of large amounts of this naphtha to produce SNG will only result in increasing both the importation and the cost of refined petroleum products, such as naphtha. It has been suggested that only those oil-based SNG plants already proposed would require additional oil imports of 646,000 barrels (bbl) per day (22). Although the actual cost for SNG from naphtha has not been published, the resulting gas would be expected to cost in the range of $\$2.50$-$4.00/kft^3$.

The capital investment (1976 dollars) for a nominal 250 Mft^3/day SNG-from-coal plant ranges from \$870 M to \$1.5 G (depending on the process involved). When compared with the current total U.S. consumption of 20 Tft^3, the output of one of these coal-to-gas plants represents only about 0.4% of this total annual demand for natural gas. In addition, none of these gasification systems have been constructed or tested on a commercial scale in the United States as yet. Large-scale production of pipeline quality gas from coal will not become an important energy source until the 1990's, although one or two plants are expected onstream by the late 1980's.

The projected gas prices from commercial coal-to-gas plants depend on the process selected, the coal to be gasified, the accounting procedures used, and the source being cited. These prices for SNG from coal have recently been estimated (29) to range from \$2.70 to $\$6.72/kft^3$. Again it should be pointed out that these costs are based on small pilot-plant operations (<2.4 Mft^3/day) and are subject to considerable revision as more development work is completed.

Importation of Natural Gas From Canada

After an initial period of net U.S. exports to Canada in the early 1950's, the United States increased its imports of natural gas from 1958 until reaching a peak of 1.0 Tft^3 in 1973 and then slowly declined to 0.95 Tft^3 in 1976 (see Figure 21). These imports were brought about by the discovery of large natural gas reserves in Western Canada and the lack of pipelines to transport this gas to consumers in Eastern Canada. As a result, excess Western Canadian gas was sent across the border to consumers in the upper Midwest while American gas from Texas and Louisiana was pumped to Eastern Canada through transcontinental pipelines. However, after the oil crisis of 1973, the amount of gas exported to the United States was decreased. The price for Canadian gas delivered to the border has increased during the past 4 yr, from $\$0.40/kft^3$ in 1973 to about $\$1.20/kft3$ in 1977 and these increases are expected to continue until

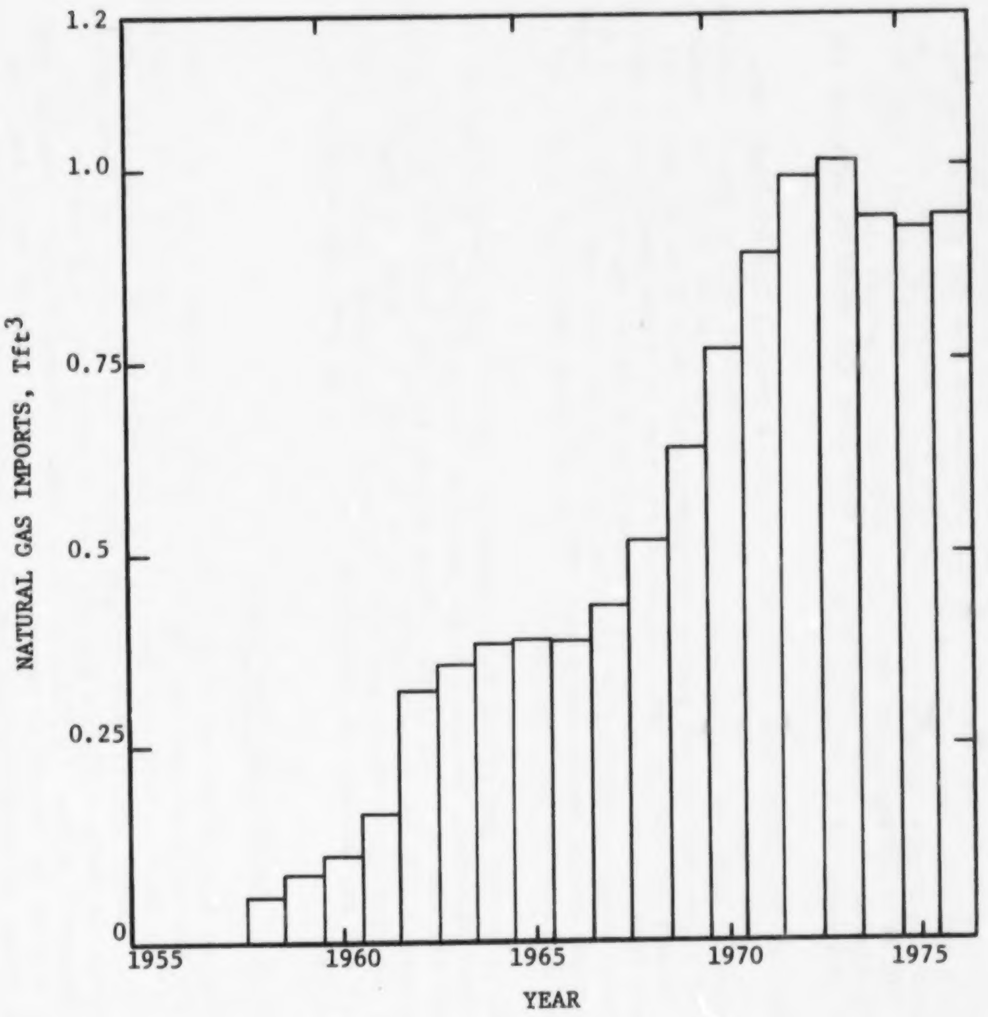

Figure 21. Natural gas imports from Canada during the period 1955-1976 (34).

the cost of exported gas is equivalent to that of imported oil. The price of this gas has recently been projected to rise to about \$2.70/kft^3 by 1980 (46).

Importation of Natural Gas From Mexico (6, 15, 16, 18)

Although imports by pipeline from Canada are expected to decline in the near future due to increased demand within Canada and the possible construction of a trans-Canadian pipeline to transport the gas to eastern provinces, the prospects of importing natural gas from Mexico have brightened considerably since the early 1970's. During the late 1950's and continuing through the 1960's Mexico exported relatively small amounts (0.05 Tft^3) of natural gas to the United States (see Figure 22), but in the 1970's this gas was needed within Northern Mexico and the exports were stopped. However, many of the wells in the recently discovered Reforma field in Southern Mexico are producing six or seven times as much gas as orginally expected and the Mexican oil company, PEMEX, is faced with a substantial surplus of natural gas which must be either consumed or flared. PEMEX had orginally planned to construct seven large NH_3 plants with a total capacity of 3 Mtons/yr; even so, considerable natural gas, which otherwise would be flared, will be available for export. Due to other economic considerations which eliminate the potential use of this gas within Mexico, exploratory talks have recently begun on the construction of a 750-mi gas pipeline from the Reforma area of Mexico to the Texas border at Reynosa and the export of 1-2 Gft^3/day of gas to the United States beginning in 1979 (35).

Even though this 2 Gft^3/day appears to be a significant amount of gas, it represents only about 3.5% of the current U.S. natural gas supply. On the other hand, this 2 Gft^3 of natural gas per day, if applied solely for NH_3 production would generate about 63,500 tons/day or 20,950,000 tons/yr, i.e., more than the total U.S. NH_3 consumption at the present time. The selling price of this gas, although not published at the present time, is expected to be equivalent to the projected price of heating oil in New York harbor or about \$2.50/kft^3 delivered at the U.S.-Mexican border (6, 18).

Although this project is only tentative at the present time, the combination of a huge surplus of natural gas in the Reforma area of Mexico (estimated 1980 production, 10 Gft^3/day), the need by PEMEX to generate revenues to finance other petrochemical plants (projected natural gas treatment capacity in 1980, 2.34 Gft^3/day), and the shortage of natural gas in the United States all seem to indicate that this pipeline will be completed and in operation supplying an initial 1.0 Gft^3/day of badly needed gas to the United States. This could be possible by as early as 1979 or 1980 and reaching full capacity of 2.0 Gft^3/day by 1981.

Even with this much gas being withdrawn and pumped to the United States, sufficient gas would still be available for constructing conventional NH_3 plants within Mexico. PEMEX has already contracted for four of these large NH_3 plants consuming about 0.189 Gft^3/day and producing 6000 tons/day (about 2 Mtons/yr) of NH_3 and plans are underway for at least three more large NH_3 plants. These four contracted plants will almost double the current Mexican NH_3 capacity to 13,000 tons/day. Additional NH_3 plants, financed by American companies, could be built and the resulting NH_3 shipped to Gulf ports by barging across the Gulf of Mexico.

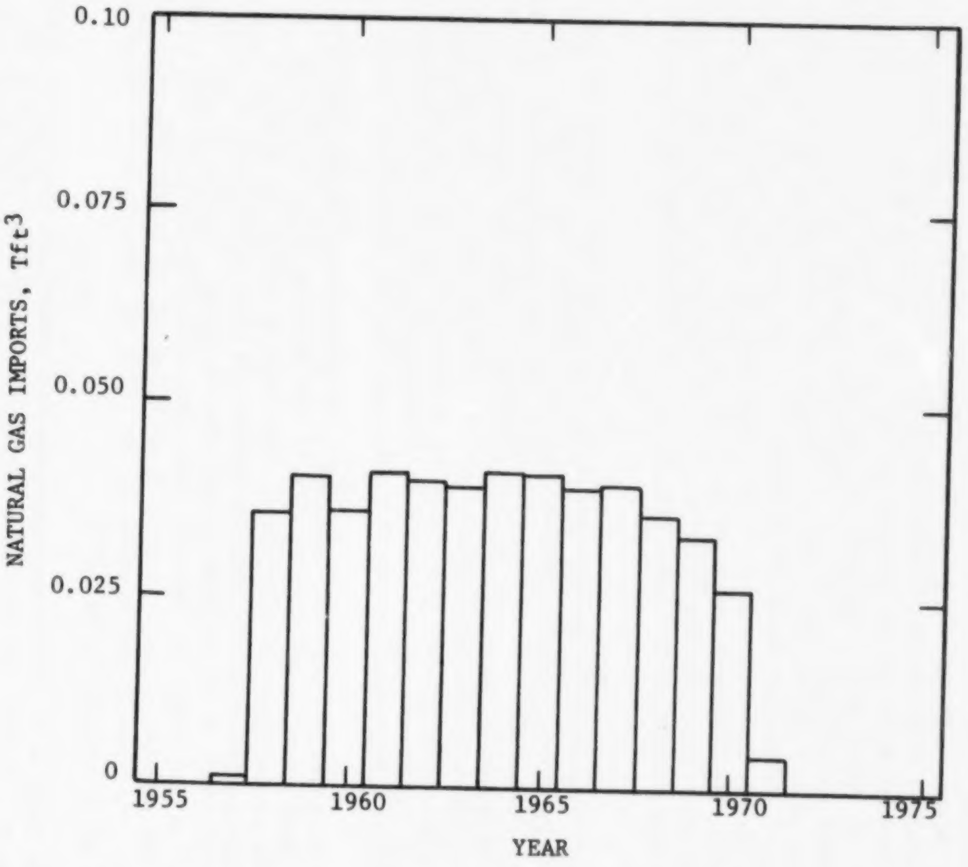

Figure 22. Natural gas imports from Mexico during the period 1955-1976 (34).

Alaskan Natural Gas (3, 17, 21)

With the discovery of large oil and associated natural gas deposits on the Alaskan North Slope and the recent completion of the trans-Alaskan oil pipeline, various proposals have been advanced to bring this gas to the contiguous 48 states rather than either reinjecting or flaring the gas. Although this is not the first major natural gas discovery in Alaska (reserves of about 5 Tft^3 were discovered earlier at Kenai), it is by far the largest with probably reserves of about 26 Tft^3.

At the present time, the favored proposal for bringing this North Slope gas south is the so-called Alcan route, a 2754-mi pipeline from Prudhoe Bay south along the trans-Alaskan oil pipeline to near Fairbanks and then southeast along the Alaskan highway to the U.S.-Canadian border. Portions of this system would consist partly of existing pipelines with about 1800 mi of new pipe needed. The initial capacity of the pipeline would be 2.6 Gft^3/day (about 1.0 Tft^3 annually) and cost between \$6.7-10.0 G (3, 25). Assuming a well-head price of \$1.00/kft^3, the estimated cost of the gas at the U.S.-Canadian border would range, depending on the source, from \$1.79-3.50/kft^3 (3, 17, 21). With construction scheduled to begin in 1979, the first gas probably would not be available before 1982 and potential delays in initiating construction of up to 2 yr are already being mentioned.

Importation of Liquefied Natural Gas (10)

Although proposals for transocean shipping of natural gas have been discussed for the past decade, the importation of LNG is only slightly closer to commercial realization in the United States today than it was 10 yr ago. In gas-starved nations such as Japan, the large-scale importation of LNG is a commercial reality but in the United States, which has substantial natural gas reserves, the economics did not and still do not justify the large-scale use of LNG. Its primary use in the United States at the present time is limited to a role as a peak-shaving energy source during periods of high demand. However, by 1985 LNG imports are expected to contribute substantial amounts of natural gas to supplement declining domestic gas production.

There are only a few countries in the world (shown in Table 51) with large natural gas reserves and low domestic consumption to guarantee a good, long-term (20-25 yr) supply of natural gas. Since the most commonly discussed standard LNG plant has a 1 Gft^3/day capacity, minimum proven reserves of about 10 Tft^3 are needed to support the construction of a natural gas liquefication plant. Of the countries listed in Table 51 as potential exporters of LNG, only Malaysia and Australia with relatively small domestic reserves actually have a large potential as a noncontracted source of LNG for the United States during the 1980's, as shown in Table 52. Russia with perhaps the largest reserves has contracted to sell some of its excess gas to Eastern and Western Europe (5.1 Gft^3/day) and will probably sell additional gas to Japan in the near future. Russian exports of LNG to the United States, although greatly discussed do not appear promising at the present time. Iran with the second largest reserves has decided that internally consuming the available gas is in its national interest and hence little additional gas will be available for export until the mid-1980's. Algeria has already contracted its available natural gas for LNG export and Saudi Arabia has only recently begun construction of a gas-gathering system. Thus most of the potential gas available for LNG already has been or will be contracted for in the near future. Additional supplies of LNG may not be available.

TABLE 51. NATURAL GAS RESERVES OF VARIOUS COUNTRIES (49)

Country	Natural gas reserves, Tft^3	
	Proven	Potential
USSR	850	2,000
Iran	330	750
Algeria	125	150
Saudi Arabia	107	-[a]
Nigeria	45	90
Abu Dhabi[b]	33	-
Indonesia[c]	31	-
Australia	30	-
Iraq	27	-
Libya	26	-
Malaysia	10	-
Qatar	8	-

a. Unknown.
b. Includes Bahrein, Dubai, Oman, and UAE.
c. Includes Arum, Kalimantan, Brunei, and Papua.

Eight base-load LNG plants (i.e., gathering, treating, liquefying, and exporting) are now operating worldwide and eight more are in various stages of construction. These plants and various applicable information about each plant are listed in Table 53. Using the announced and probable contracts for these plants, the expected world trade in LNG for various years is shown in Table 54. From this table it is apparent that the United States and Japan are the two largest potential buyers, importing 49% and 28% of the world supply of LNG in 1985, respectively, while Algeria and Indonesia are expected to export nearly two-thirds of the available LNG.

It must also be kept in mind that a typical LNG tanker carries the equivalent of 2700 Mft^3 of gas. If one round trip requires about 24 days, a single ship could supply only about 0.041 Tft^3 annually or about 0.2% of the current annual U.S. natural gas demand. With the cost of a single LNG tanker currently estimated at $120-140 M, another limiting factor in the worldwide LNG trade could be the availability of LNG tankers. Thus, although LNG imports are expected to provide a substantial amount of supplemental natural gas in the future, LNG alone will not solve the present natural gas shortage.

TABLE 52. LNG EXPORTING COUNTRIES IN 1985 (49)

Country	Natural gas production, Tft^3/yr	Potential gas surplus, Tft^3/yr	Contracted gas exports (country), Tft^3/yr	Noncontracted gas available for LNG, Tft^3/yr
USSR	19.0	3.4[a]	1.9 (Europe)	1.5
Iran	5.6	1.1	0.4 (Europe) 0.35 (USSR) 0.35 (USA and Japan)	0.0
Algeria	2.5	2.1	0.6 (Europe 1.5 (USA)	0.0
Saudi Arabia	1.6	0.0	-	0.0
Nigeria	0.3	0.29	0.29 (USA)	0.0
Abu Dhabi	1.0	0.5	0.5 (Japan)	0.0
Iraq	0.4	0.07	-	0.07
Qatar	0.5	0.3	-	0.3
Malaysia	0.4	0.36	-	0.36
Indonesia	0.7	0.65	0.5 (Japan) 0.15 (USA)	0.0
Australia	1.3	0.36	-	0.36

a. Includes imports of 0.4 Tft^3/yr.

TABLE 53. CURRENT AND PROPOSED LNG BASE-LOAD PLANTS (50)

Site	Yr of operation	LNG plant capacity, Mft^3/day	Importing country (amount, Mft^3/day)
Current			
Algeria			
Arzew	1964	200	France (50) Spain (40) Great Britain (100)
Skikda	1972	450	France (350) U.S. (44)
Indonesia			
Brunei	1974	750	Japan (741)
Libya			
Marsa El Berga	1970	385	Italy (235) Spain (110)
U.S.			
Kenai, Alaska	1969	173	Japan (160)
Committed			
Algeria			
Arzew	1977[a]	1,100	U.S. (1,000)
Skikda	1975	170	U.S. (61.4)
	1976[a]	350	U.S. (600) Spain (436)
Indonesia			
Kalimantan	1977[a]	1,200	Japan (1,050)
Abu Dhabi			
Das Island	1976	350	Japan (450)
Probable			
Algeria			
Arzew	1979[a]	1,000	Belgium (340) U.S. (1,345)
Chile			
Cabo Negro	1979[b]	290	Brazil (220)
Indonesia			
Arun	1977[a]	1,220	Japan (550)
Malaysia			
Sarawak	–[c]	–	Japan (750)
Nigeria			
Conney	–[c]	800	U.S. (800)
U.S.			
Cook Inlet	1978[b]	400	U.S. (400)

a. Under construction.
b. Planned.
c. Unknown.

TABLE 54. CURRENT AND PROJECTED LNG IMPORTS AND LNG EXPORTS BY COUNTRY (55)

	LNG trade, Mft^3/day			
	1977	1980	1983	1985
Importing country				
U.S.	42	1,816	4,570	6,200
W. Europe	986	1,625	2,950	2,950
Japan	1,180	1,605	3,480	3,480
Total	2,208	5,046	11,000	12,630
Exporting country				
Algeria	693	2,976	5,845	6,675
Libya	335	335	335	335
Nigeria	-	-	-	800
Iran	-	-	800	800
Alaska	135	265	535	535
Indonesia	750	1,175	2,340	2,340
Abu Dhabi	295	295	295	295
Malaysia	-	-	850	850
Total	2,208	5,046	11,000	12,630

The expected price of this imported LNG has also increased substantially in the past decade. Although originally sold at near production cost, with the Arab oil embargo and the quadrupling of oil prices, the selling price of LNG in the exporting countries has also substantially increased to reflect the higher demand for premium fuel. The projected selling price of imported LNG in the United States is currently estimated to range from \$3.00-3.50/kft^3, which is essentially double the current price of conventional domestic natural gas. Future prices for delivered LNG are expected to be equivalent (on a \$/MBtu basis) to that of imported crude oil, which has been estimated at about \$3.67 in 1980 and \$4.67/MBtu in 1985 (10). In fact a recent cost for Algerian gas to be delivered to the United States in 1984 was estimated to be about \$4.57/kft^3 (10).

New Technologies (12, 41)

Other potential future sources of natural gas include various new technologies such as methane from coal mines, gas from tight shales and methane from geopressurized brine solutions. Of these new technologies that is closest to commercial realization is the recovery of methane associated with coal seams. One of the major problems associated with mining coal is the potentially explosive nature of methane in coal mines. The flammability limits of methane range from approximately 5 to 15% by volume in air and in the past numerous mine accidents were caused by methane explosions. By operating the mine with enough ventilation such that the methane concentration in the mine remains below the 5% explosive limit, coal mining has become a much safer occupation. This resulting methane-in-air mixture is swept from the mine and exhausted to the atmosphere.

This large-scale waste of a valuable energy source, although necessary in the past with the low cost and ready availability of natural gas, is now undergoing increased attention. For example, the FPC has recently proposed exempting this methane from Federal regulation and thus give mines an economic incentive to sell this gas at whatever price the consumers are willing to pay. Although minor amounts are already being sold to interstate pipelines, it has been estimated that about 250 Mft^3/day of methane is either being vented or flared from currently operating U.S. coal mines.

This methane was apparently formed as the organic material, orginally laid down within the seam, gradually decomposed under heat and pressure to form coal. The amount of methane remaining in the seam depends on the permeability of the underlying and overlying rock strata and also the amount of fracturing in the coal. Some particular seams contain large amounts of methane while others contain very little.

ERDA has estimated that up to 749 Tft^3 of methane may be trapped in various coal seams and about 30% of this gas could be recovered. Although no method for removing this methane has been tested on a commercial scale as yet, it has been suggested that by drilling through a coal seam and then fracturing the coal by explosives could release the methane and allow its removal from the mine. No costs have been estimated for this gas and additional development work is needed before the large-scale application of this technology will be available.

Estimates of the natural gas reserves held in eastern tight shales range up to 1200 Tft^3, a potential 60-yr supply at current rates of consumption. Although this natural gas deposit stretching from New York and Michigan south to Alabama has been extensively tapped during the past 50 yr, numerous problems have prevented its widespread use as a source of natural gas. The largest drawback associated with these deposits is that the shale in which the gas is trapped is relatively impermeable. Unless the well happens to hit a natural fracture, gas production is very slow and may not economically justify a gas-collection system.

Potential recovery techniques must include some method of causing large fractures in the tight shale formations and at the present time this is the limiting technical factor in removing the gas. Various options such as the injection of water under pressure, conventional explosives, and even nuclear explosions have been used in the past with limited success. However, as long as the price of natural gas remains low, natural gas production from these shale formations will not provide significant amounts of gas.

The other potential source of gas listed under new technologies is the recovery of natural gas from geopressurized brine solutions. These hot brine solutions containing up to 60 ft^3 of dissolved methane per barrel of brine are located at depths of 10,000 to 20,000 ft over an area estimated at 80,000 to 160,000 sq mi along the Texas and Louisiana coast. A recent estimate by the USGS has put the total recoverable reserves of methane at 490 Tft^3.

Numerous methods of recovering both the methane and the inherent energy of the hot, pressurized brine solutions have been suggested but the most promising appears to be a stepwise reduction of the pressure, releasing the methane and turning a power recovery turbine. A plant based on this method and treating 50,000 bbl/day of brine would produce about 8 MW of electricity in addition to 3 Mft^3 of methane per day (assuming 60 ft^3/bbl). However, substantial development work remains to be done both in confirming various optimistic assumptions and obtaining more data on these geopressurized zones. Some of these major problem areas include: (1) permeability of the zones and hence productivity of the region, (2) physical properties of the brines (primarily methane content), and (3) feasibility of spent brine reinjection.

Other Sources of Natural Gas

Although technically not an additional source of natural gas, several recent developments have suggested ways by which additional supplies of natural gas may be freed for other uses in the future. The first involves two potential innovations in the NH_3 plant which would reduce the amount of natural gas consumed per ton of product while the second pertains to the redirection of scarce natural gas away from simple burning of gas as a boiler fuel.

Of the two process modifications to the conventional NH_3 plant currently undergoing research and development, the first and potentially most important modification is the recent work done on replacing natural gas as the fuel used to indirectly heat the primary reformer, with vaporized fuel oil. About one-third of the natural gas consumed in producing NH_3 (12,600 ft^3/ton) is effectively wasted in heating the primary reformer. By substituting another fuel, such as vaporized fuel oil, in all natural gas-based NH_3 plants in the future, the natural gas required for producing NH_3 could be reduced by about 40%. The second modification (24) involves adding a cryogenic separation step to the purge gas stream from the synthesis loop. Since nearly 65% of this purge gas is H_2, using a cryogenic separator to recover 90% of this H_2 and recycling it to the synthesis loop has been estimated to yield an additional 5% NH_3 for the same input of natural gas (savings--approximately 1500 ft^3/ton of NH_3).

In addition to these potential natural gas savings within the NH_3 industry, the realization of the value of natural gas has led others to project that natural gas will be redirected toward those uses where natural gas must be used exclusively. For example, FEA has recently begun to prohibit the use of natural gas as a primary energy source for both utility and industrial users. Since the consumption of natural gas in utility boilers alone is currently estimated (44) at about 3.5 Tft^3 annually (about 17.5% of the current supply), most of this gas will gradually become available for other uses as more of the older plants are retired and no new gas-fired power plants are built. In fact, it was recently estimated (25, 50) that natural gas consumption in utility boilers would decline to 1.5 Tft^3 by 1985 and to less than 1.0 Tft^3 in 2000. Although no estimate of gas savings from the phaseout of gas use in industrial boilers is available, from the utility boiler savings alone it is apparent that significant amounts of gas can be freed for other uses, especially the production of NH_3. This is not meant to imply that this additional gas will be used exclusively or even partially to produce NH_3, only that the available gas may be directed to higher priority uses.

CONCLUSIONS

Various processes for removing NO_x from boiler flue gas are now undergoing development and one of the more promising and the furthest developed method is the SCR-type system. SCR processes involve the injection of NH_3 into the flue gas ducts and passing the resulting gas mixture over a base-metal catalyst. The NO_x and the NH_3 react to form N_2 and H_2O which are then exhausted. The entire system is relatively simple with only one major piece of equipment, the catalytic reduction reactor, and no major byproduct streams requiring further treatment.

The most critical question associated with wide-scale commercial application of this NO_x control technology has been, "What will the impact on NH_3 production and marketing be?" The NH_3 is only injected at a rate such that the mol ratio of NH_3:NO_x is about 1:1 and the NO_x concentration is very low (i.e., about 600 ppm for coal-fired flue gas). However, the increasing emphasis both on using coal as a boiler fuel and on using increased amounts of electrical energy has led others to project the possibility of severe economic disruptions in the NH_3 market in the future if NO_x FGT is required for all large boilers.

ONSITE GENERATION VERSUS ONSITE STORAGE OF NH_3 FOR NO_x FGT SYSTEMS

The annual NH_3 consumption for an NO_x FGT system for typical 500-MW coal- and oil-fired boilers range from 6266 tons/yr for the 500-MW coal-fired boiler without combustion modifications to 772 tons/yr for a 500-MW oil-fired boiler with combustion modifications. For a plant to produce NH_3 at a competitive price it must have a minimum capacity of approximately 1000 tons/day (330,000 tons/yr) and clearly even a 2000-MW coal-fired power plant, which is a large unit by present standards, could not economically justify a captive NH_3 plant. Instead the NH_3 would be purchased from an independent producer, shipped to the power plant, and stored onsite until needed.

IMPACT OF NH_3 DEMAND FOR NO_x FGT ON THE DOMESTIC NH_3 MARKET

As a basic premise for calculating the annual NH_3 consumption for NO_x FGT, it was assumed that only new fossil fuel-fired boilers coming online after January 1, 1985, would be required to install 90% efficient NO_x control systems. The capacity of new utility boilers coming online each year and the fuel mix in these new boilers during the period 1985-2000 was then estimated by using the recent projection of the National Electric Reliability Council as the base case for both the fuel mix and the total domestic generating capacity in 1985. It was further assumed that the total U.S. generating capacity would double during the period 1985-2000 from 798 GW to about 1660 GW. Assuming that no new gas-fired and very few oil-fired utility boilers will be brought online after 1985 and that the current problems with bringing nuclear units online will continue, coal would be the fuel for 50% (829 GW) of all the utility boilers generating electricity in 2000.

The annual coal-fired electrical capacity additions needed to meet this projected coal-fired generating capacity in 2000 steadily increased from the 15 GW added in 1985 to 55 GW in 2000. The resulting annual NH_3 requirement for NO_x FGT, assuming the average 500-MW coal-fired boiler emits 600 ppm NO_x (3000 lb of NO_x/hr) in the flue gas, ranges from 0.09 Mtons in 1985 to 6.7 Mtons in the year 2000 for a 90% efficient SCR system and from 0.04 Mtons in 1985 to 2.9 Mtons in the year 2000 for the case of combustion modification followed by an 80% efficient SCR system. For case 1 (90% efficient SCR system) this annual NH_3 requirement NO_x FGT ranges from 0.44% of the total projected domestic NH_3 demand in 1985 to 20.1% in the year 2000. For case 2 (combustion modification followed by an 80% efficient SCR system) these percentages are 0.19% and 8.7% respectively. The requirement for 90% NO_x removal from utility boiler stack gas using only SCR will require the addition of about one new 1000 ton/day NH_3 plant each year while using combustion modifications with an 80% efficient SCR system will dictate about one new 1000 ton/day NH_3 plant every 2 yr. Thus the requirement for 90% NO_x removal as outlined in the premises of this study will probably not have a severe impact on the NH_3 market. Instead of the projected 3% annual increase in NH_3 demand during the period 1978-2000, the additional consumption of NH_3 for NO_x FGT will result in an annual 4.5% increase in NH_3 demand during the years 1985-2000.

In addition to large utility boilers, large industrial boilers contribute a significant fraction of the total amount of NO_x released in the United States and hence may become subject to the same requirement for 90% NO_x removal from all new sources previously postulated for large utility boilers. Although the total capacity of all large U.S. industrial boilers has not been published, this industrial boiler capacity was estimated as about 25% of the utility boiler capacity. Using the same emission assumptions as was used for the utility boilers, the annual NH_3 consumption for NO_x FGT controls on all new large boilers (both utility and industrial) would range from 0.12 Mtons in 1985 to 8.4 Mtons in the year 2000. Although these NH_3 usage rates are higher than those for new large utility boilers only, the increase is not significant enough to cause any further dramatic increase in demand for NH_3.

If, on the other hand, the premises had stated that all large boilers, i.e., old and new, would require 90% efficient NO_x control the result would have been substantially different. Under this assumption, the estimated NH_3 demand for NO_x control would range from 4.0 Mtons in 1985 (19% of projected NH_3 demand) to 10.4 Mtons in 2000 (31.4% of projected NH_3 demand). Such an increase in demand in 1985 would most likely result in a large increase in the price of NH_3.

The major impact of the increased NH_3 consumption due to the adoption of more stringent NO_x controls may not necessarily be on the domestic NH_3 market but rather the indirect effect on the availability and price of natural gas. With the current national energy policy geared toward reducing our dependence on natural gas and the current uncertainty concerning the future availability of natural gas, even these small amounts (relative to current gas production) may in the future be considered a significant impact.

CURRENT NH_3 PRODUCTION ECONOMICS

Although numerous potential H_2 feedstocks are available for the production of NH_3, approximately 95% of the NH_3 produced domestically is made from natural gas. The remaining 5% of current production comes from various miscellaneous sources such as off-gases from electrolytic cells and coke-oven gas. However, these other sources are usually unique systems in which various factors have come together to provide an economic incentive for NH_3 generation. The potential of these sources for producing a larger share of the domestic NH_3 supply is not promising.

At the present time, the widespread curtailment of natural gas during the peak heating season and the steady decline in both proven reserves and production of conventional natural gas have led to concerns that enough natural gas will not be available to meet this additional demand. Conventional natural gas supplies and reserves are declining at the present time but supplemental supplies of natural gas, such as LNG, SNG (from oil and coal), Alaskan gas, and imports from Canada and Mexico, are increasing. The cost of this natural gas, however, is expected to be sharply higher than current prices ($1.45-2.00/MBtu) which are themselves significantly higher than prices of only 2-3 yr ago.

These higher natural gas prices and questions about the future availability of natural gas have revived interest in the use of alternative feedstocks for the production of NH_3. Any hydrocarbon is a potential H_2 feedstock and since oil and coal are widely availabe these are the two most commonly mentioned alternative feedstocks. Various processes for converting either oil or coal into H_2 to be used for NH_3 synthesis are technically feasible and are commercially available on the world market. The major stumbling block for domestic acceptability is the process economics.

Each of these alternative feedstocks requires a higher initial capital investment and thus at the same fuel cost, the unit cost of producing NH_3 would be higher for both naphtha and coal than for natural gas. Thus for either naphtha or coal to compete economically with natural gas, this disadvantage must be offset by a significant feedstock cost differential.

At the present time and under the premises of this study, the feedstock cost differential is not large enough for alternate feedstocks to economically compete with natural gas since, even at a natural gas price of $2.00/MBtu, NH_3 can be produced for about $143/Mton while current naphtha and coal costs yield NH_3 production costs of $193 and $186/Mton respectively. However, with the cost of natural gas expected to increase in the near future, this feedstock cost differential between natural gas and alternate feedstocks may increase sufficiently to justify the production of NH_3 from these other feedstocks.

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